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Volume II. Gaseous Emissions
Solid Wastes

Pullman Kellogg, Houston, TX

Prepared for

Industrial Environmental Research Lab, Research Triangle Park, NC

Oct 79

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Environmental Protection
Agency

Industrial Environmental Research
Laboratory
Research Triangle Park NC 27711

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October 1979

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EPA

Coal Conversion Control Technology Volume II. Gaseous Emissions; Solid Wastes

Interagency Energy/Environment R&D Program Report

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Coal Gasification	Stationary Sources	13H
Coal Preparation	Coal Conversion	13I
Wastes	Solid Wastes	

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Volume II. Gaseous Emissions; Solid Wastes

by

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Contract No. 68-02-2198
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Prepared for

U.S. ENVIRONMENTAL PROTECTION AGENCY
Office of Research and Development
Washington, DC 20460

i.a

ABSTRACT

Information has been gathered on coal conversion process streams. Available and developing control technology has been evaluated in view of the requirements of present and proposed federal, state, regional and international environmental standards. The study indicates that it appears possible to evolve technology to reduce each of the components of each process stream to an environmentally acceptable level. The conclusion has also been reached that such an approach would be costly and difficult of execution.

Because all coal conversion processes are net users of water, liquid effluents need be treated only for recycling within the process, thus achieving essentially "zero discharge." Further, with available technology gaseous emissions can be controlled to meet present environmental standards, particulates can be controlled or eliminated and disposal of solid wastes can be managed to avoid deleterious environmental effects.

Volume I focuses on environmental regulations for gaseous, liquid, and solid wastes, and the control technology for liquid effluents. Volume II deals with the control technology of gaseous emissions and solid wastes.

Volume III includes a program for economic analysis of control technology and includes the appendix.

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SECTION 9

ENVIRONMENTAL DATA ACQUISITION: CONTROL OF GASEOUS EMISSIONS

DEVELOPMENT OF CONVERSION PROCESS EMISSION STREAM MODELS

The study of available information on gaseous emissions from coal conversion processes paralleled the study on liquid effluents, discussed in Section 8, and utilized the same sources. As the study progressed it became apparent that the gaseous emissions from gasification processes operating at low temperatures and those from gasification processes operating at high temperatures were, with few exceptions, basically similar and therefore solutions proposed for emission problems in one category would reasonably well fit the other. Although there were much less data and information on emissions from liquefaction processes, it appeared that the emissions from the several liquefaction processes were sufficiently similar to warrant consideration as a class, but sufficiently different from the gasification emissions to warrant separate consideration.

In the EPA planning of the overall investigations into discharges from coal conversion plant operations, other contractors were to investigate means of recovering products and byproducts from the conversion processes and determine compositions and quantities of the emission streams from the recovery operations. These emission streams were to have been the starting point for Pullman Kellogg's efforts on application of control technology of such efficiency that the final gas streams released to the atmosphere would be of a quality to meet or be better than present environmental standards.

Unfortunately, the emission stream information was not developed by the other contractors, and therefore Pullman Kellogg had no recourse but to develop process flowsheets and material balances from available published information in order to estimate the compositions and quantities of gases from recovery processes. The time required to develop the flowsheets and material balances shortened the time available, within the framework of Pullman Kellogg's contract with EPA, for investigation of emission control technology.

As a result of these constraints, the investigations take the form of general descriptions of the control technology to be applied to a gaseous emission, examples of application of the technology whenever these were available, evaluation where possible of means of increasing the control process efficiency and cost information. The results of the investigations are most valuable as a means of pointing out areas of strengths and weaknesses in available control technology and emphasizing the possibilities for future development of control technology to meet future environmental goals.

The Lurgi Dry Ash process was selected as the base case for application of control technology to gaseous emissions from coal gasification processes because of its commercial status, the amount of information available from commercial operation, the number of studies that had been made on the process, and the collaboration with the liquid effluent studies. The flowsheet in Figure 9-1 is based on the C. F. Braun conceptual designs (1)* for gasification of western (low sulfur) coal to produce 250

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*Item in reference list.

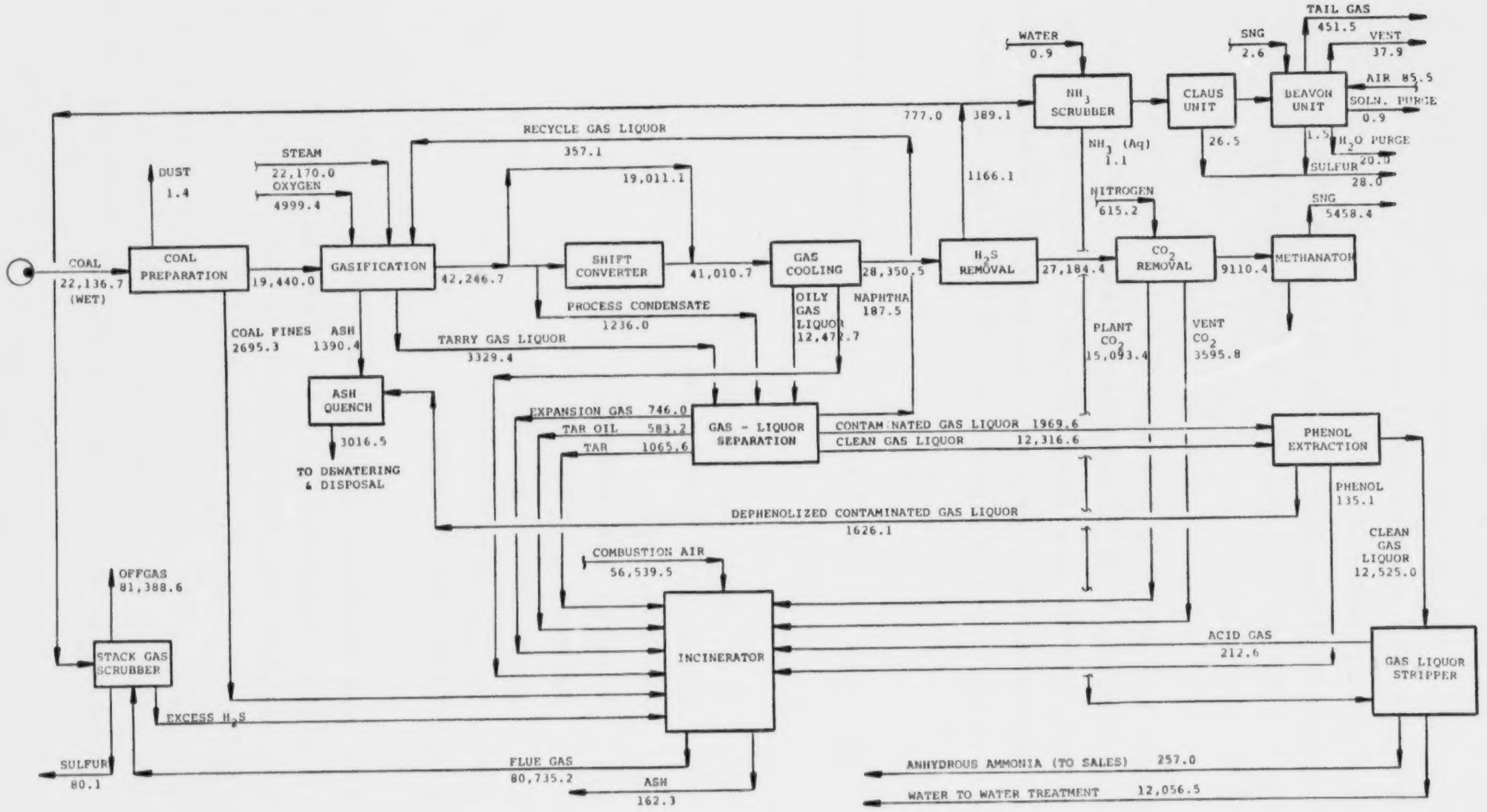

COAL
22,136.7
(WET)
COAL PREPARATION
DUST
1.4
19,440.0
COAL FINES
2695.3
STEAM
22,170.0
OXYGEN
4999.4
GASIFICATION
ASH
1390.4
ASH QUENCH
3016.5
TO DEWATERING & DISPOSAL
42,246.7
TARRY GAS LIQUOR
3329.4
RECYCLE GAS LIQUOR
357.1
19,011.1
SHIFT CONVERTER
PROCESS CONDENSATE
1236.0
41,010.7
GAS COOLING
OILY GAS LIQUOR
12,472.7
NAPHTHA
187.5
28,350.5
GAS - LIQUOR SEPARATION
EXPANSION GAS 746.0
TAR OIL 583.2
TAR 1065.6
CONTAMINATED GAS LIQUOR 1969.6
CLEAN GAS LIQUOR 12,316.6
DEPHENOLIZED CONTAMINATED GAS LIQUOR
1626.1
H_2S REMOVAL
1166.1
777.0
389.1
WATER
0.9
NH_3 SCRUBBER
NH_3 (Aq)
1.1
27,184.4
NITROGEN
615.2
CO_2 REMOVAL
PLANT CO_2
15,093.4
VENT CO_2
3595.8
9110.4
METHANATOR
SNG
5458.4
CLAUS UNIT
26.5
SNG
2.6
BEAVON UNIT
TAIL GAS
451.5
VENT
37.9
AIR 85.5
SOLN. PURGE
0.9
1.5
H_2O PURGE
20.0
SULFUR
28.0
PHENOL EXTRACTION
PHENOL
135.1
CLEAN GAS LIQUOR
12,525.0
GAS LIQUOR STRIPPER
ACID GAS
212.6
ANHYDROUS AMMONIA (TO SALES) 257.0
WATER TO WATER TREATMENT 12,056.5
COMBUSTION AIR
56,539.5
INCINERATOR
FLUE GAS
80,735.2
ASH
162.3
EXCESS H_2S
STACK GAS SCRUBBER
OFFGAS
81,388.6
SULFUR
80.1

Figure 9-1.- Flow diagram for SNG production by Lurgi gasification.

million standard cubic feet of SNG per day, supplemented in several sections with information from reports by Cameron Engineers (2), from private communications with C. F. Braun and Allied Chemical Company concerning the Selexol process and from work done by Pullman Kellogg operating within a confidentialty agreement with Allied Chemical but without disclosure of any confidential information.

Included in Figure 9-1 are a Claus plant for sulfur recovery, a Beavon plant for treatment of Claus plant tail gas and, for the incinerator/boiler, a flue gas desulfurization unit based on the U. S. Bureau of Mines citrate process. Details of these sections of the flowsheet will be discussed in "Integrated Schemes for Emission Control."

Details of the sections of the Lurgi flow diagram are shown in the following:

Figure 9-2.	Coal gasification section.
Figure 9-3.	Gas liquid treatment.
Figure 9-4	Shift conversion.
Figure 9-5.	Selexol H_2S removal.
Figure 9-6.	Selexol CO_2 removal.
Figure 9-7	Methanation and drying.
Figure 9-8	Incinerator/boiler.
Figure 9-9.	Claus-Beavon plants.

An overall feed and product weight balance for the plants is shown in Figure 9-10. Completion of the base case Lurgi process overall material balance was necessary in order to calculate the hydrogen sulfide feed to sulfur recovery for the base case for operation with low sulfur coal and for the alternate case of Lurgi operation with high sulfur coal. Development of this information then allowed inquiry into means of reducing the final

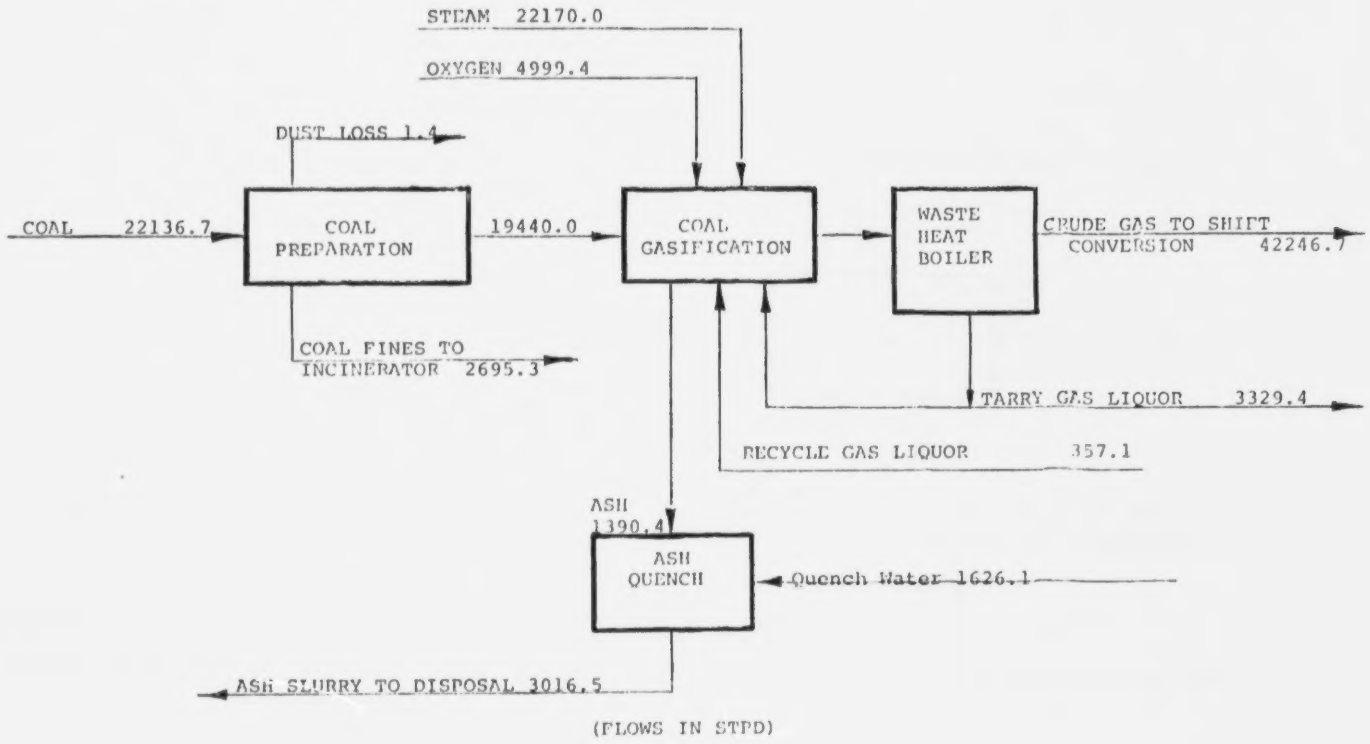
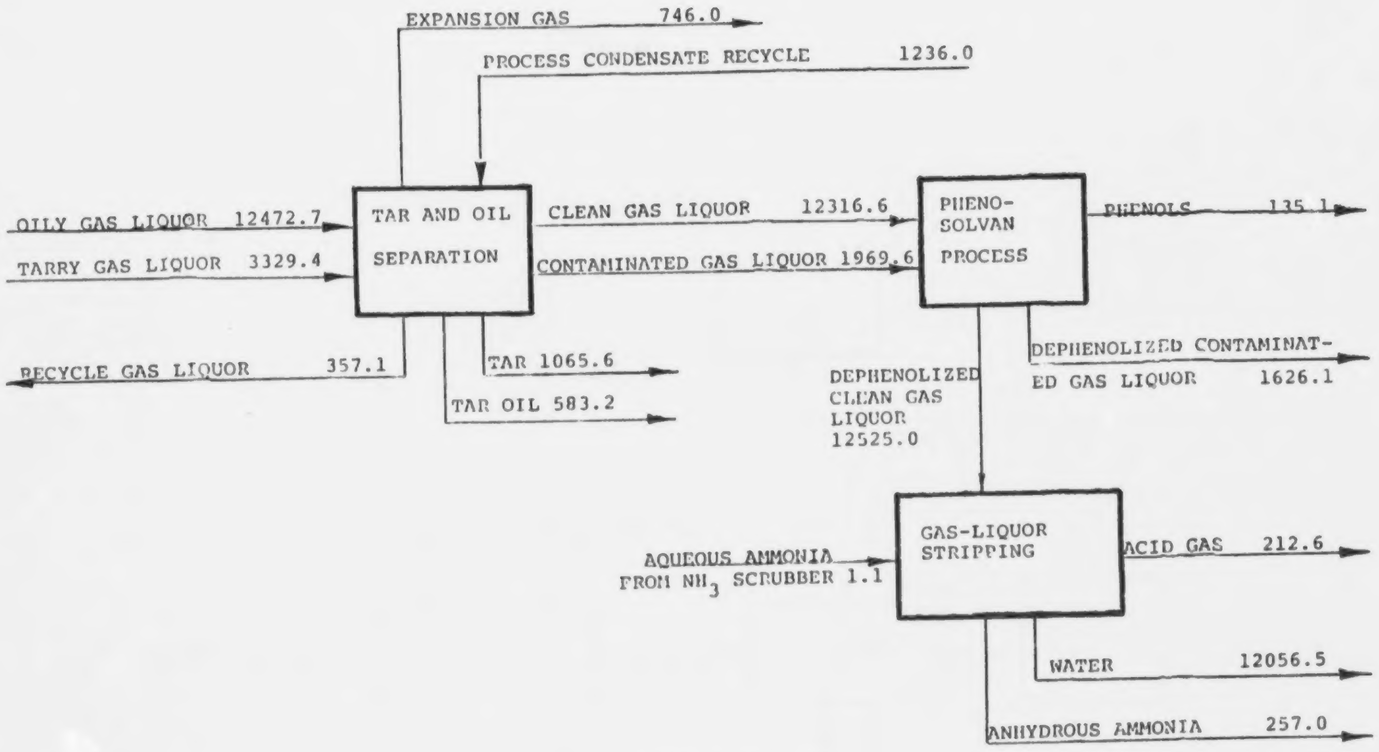

STEAM 22170.0
OXYGEN 4999.4
DUST LOSS 1.4
COAL 22136.7
COAL PREPARATION
19440.0
COAL GASIFICATION
WASTE HEAT BOILER
CRUDE GAS TO SHIFT CONVERSION 42246.7
COAL FINES TO INCINERATOR 2695.3
TARRY GAS LIQUOR 3329.4
RECYCLE GAS LIQUOR 357.1
ASH 1390.4
ASH QUENCH
Quench Water 1626.1
ASH SLURRY TO DISPOSAL 3016.5
(FLOWS IN STPD)

Figure 9-2. Lurgi process: coal gasification section.


EXPANSION GAS 746.0
PROCESS CONDENSATE RECYCLE 1236.0
OILY GAS LIQUOR 12472.7
TARRY GAS LIQUOR 3329.4
RECYCLE GAS LIQUOR 357.1
TAR AND OIL SEPARATION
TAR 1065.6
TAR OIL 583.2
CLEAN GAS LIQUOR 12316.6
CONTAMINATED GAS LIQUOR 1969.6
PHENO-SOLVAN PROCESS
PHENOLS 135.1
DEPHENOLIZED CONTAMINATED GAS LIQUOR 1626.1
DEPHENOLIZED CLEAN GAS LIQUOR 12525.0
AQUEOUS AMMONIA FROM NH3 SCRUBBER 1.1
GAS-LIQUOR STRIPPING
ACID GAS 212.6
WATER 12056.5
ANHYDROUS AMMONIA 257.0

(FLOWS IN STPD)

Figure 9-3. Lurgi process: gas liquor treatment section.

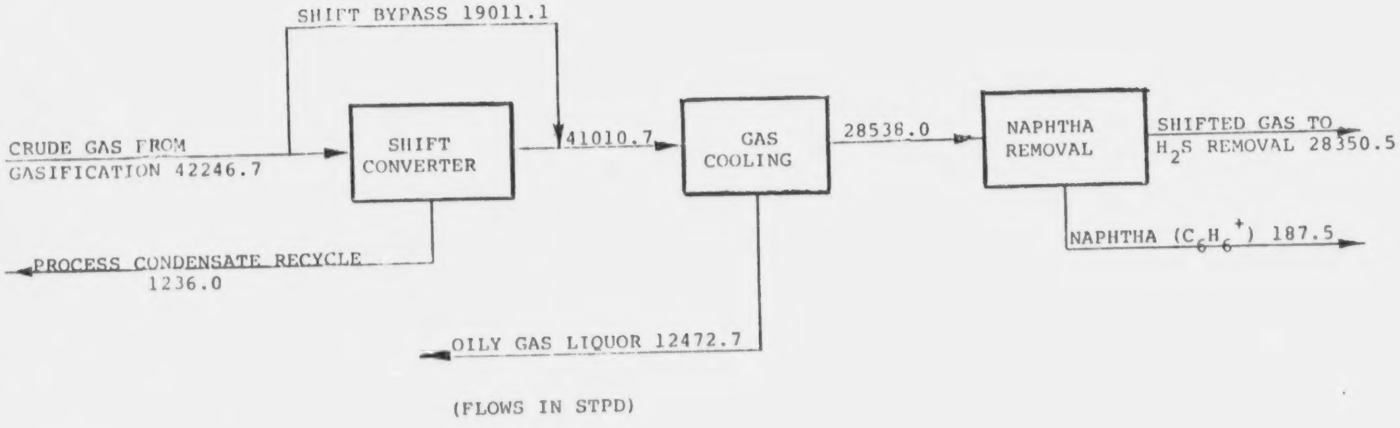
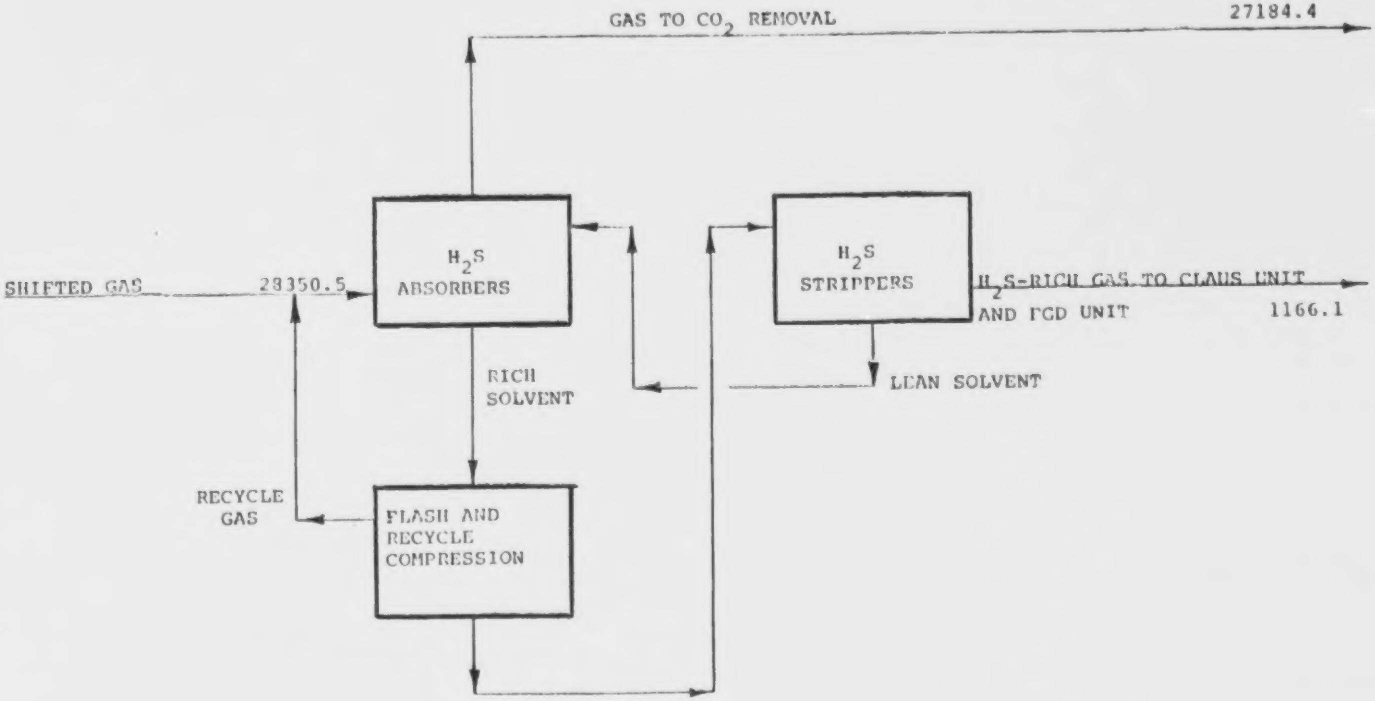

SHIFT BYPASS 19011.1
CRUDE GAS FROM GASIFICATION 42246.7
SHIFT CONVERTER
41010.7
GAS COOLING
28538.0
NAPHTHA REMOVAL
SHIFTED GAS TO H_2S REMOVAL 28350.5
NAPHTHA ($C_6H_6^+$) 187.5
PROCESS CONDENSATE RECYCLE 1236.0
OILY GAS LIQUOR 12472.7
(FLOWS IN STPD)

Figure 9-4. Lurgi process: shift conversion.


GAS TO CO$_2$ REMOVAL
27184.4
SHIFTED GAS
28350.5
H$_2$S ABSORBERS
H$_2$S STRIPPERS
H$_2$S-RICH GAS TO CLAUS UNIT AND FGD UNIT
1166.1
LEAN SOLVENT
RICH SOLVENT
RECYCLE GAS
FLASH AND RECYCLE COMPRESSION

(FLOWS IN STPD)

Figure 9-5. Lurgi process: Selexol H_2S removal.

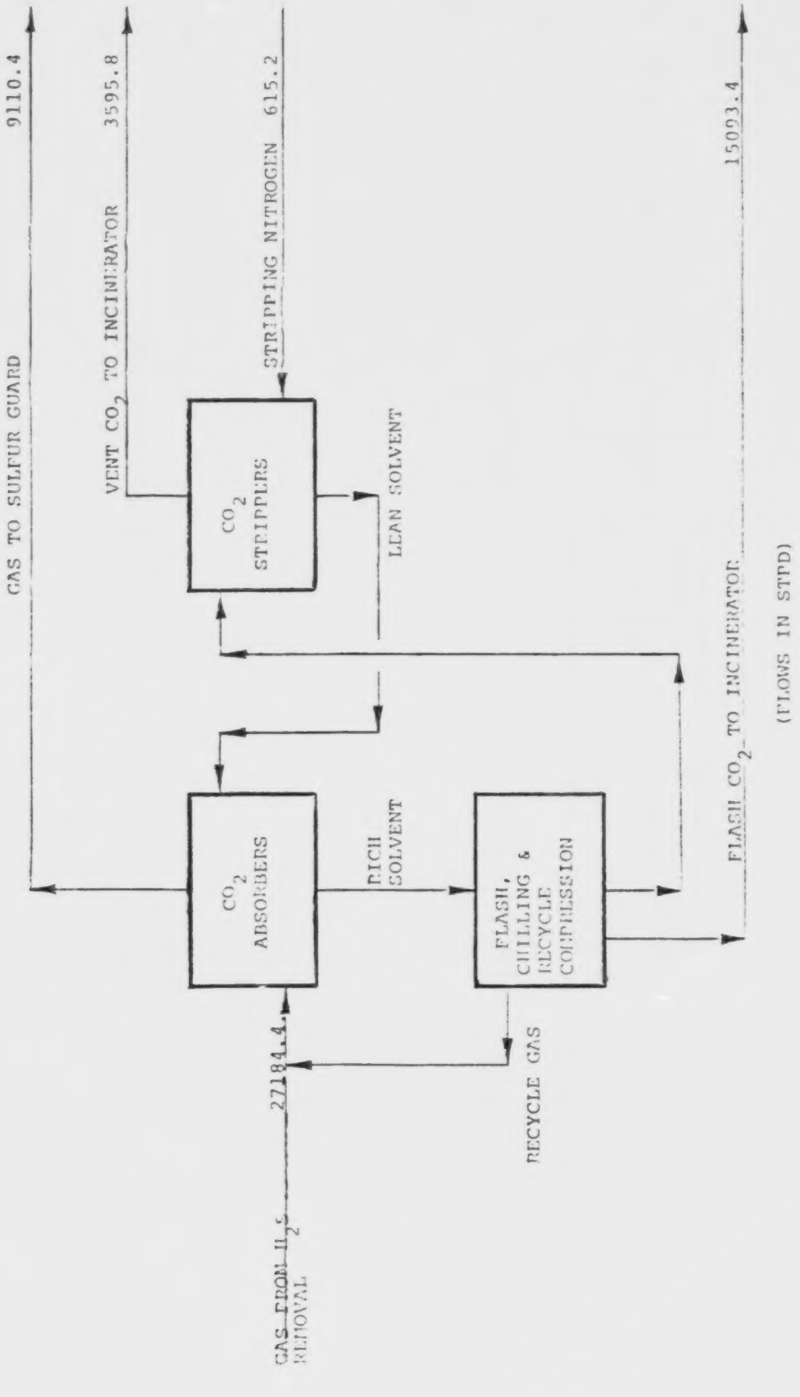
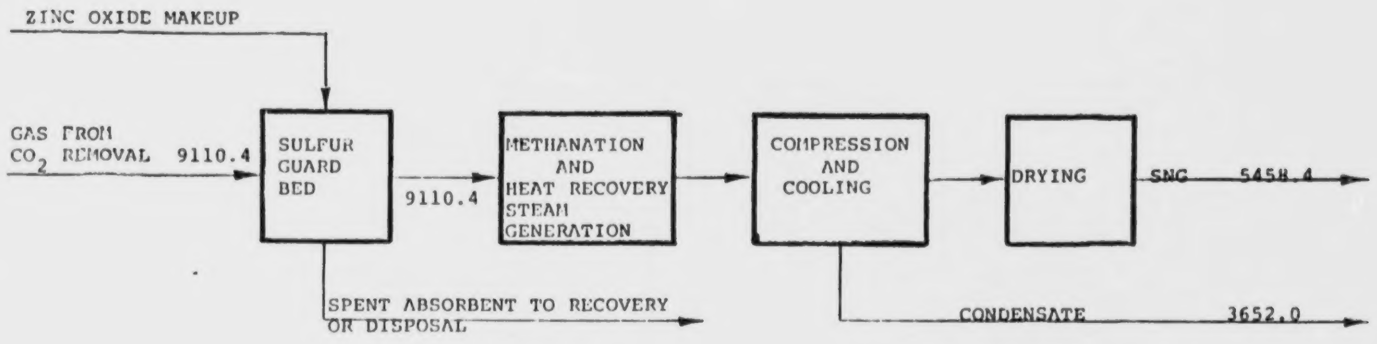

GAS TO SULFUR GUARD
9110.4
VENT CO_2 TO INCINERATOR
3595.8
STRIPPING NITROGEN
615.2
CO_2 STRIPPERS
LEAN SOLVENT
CO_2 ABSORBERS
RICH SOLVENT
FLASH, CHILLING & RECYCLE COMPRESSION
FLASH CO_2 TO INCINERATOR
15093.4
(FLOWS IN STPD)
RECYCLE GAS
27184.4
GAS FROM H_2S REMOVAL

Figure 9-6. Lurgi process: Selexol CO_2 removal.


ZINC OXIDE MAKEUP
GAS FROM CO2 REMOVAL
9110.4
SULFUR GUARD BED
9110.4
METHANATION AND HEAT RECOVERY STEAM GENERATION
COMPRESSION AND COOLING
DRYING
SNG
5458.4
SPENT ABSORBENT TO RECOVERY OR DISPOSAL
CONDENSATE
3652.0

(FLOWS IN STPD)

Figure 9-7. Lurgi process: methanation and drying

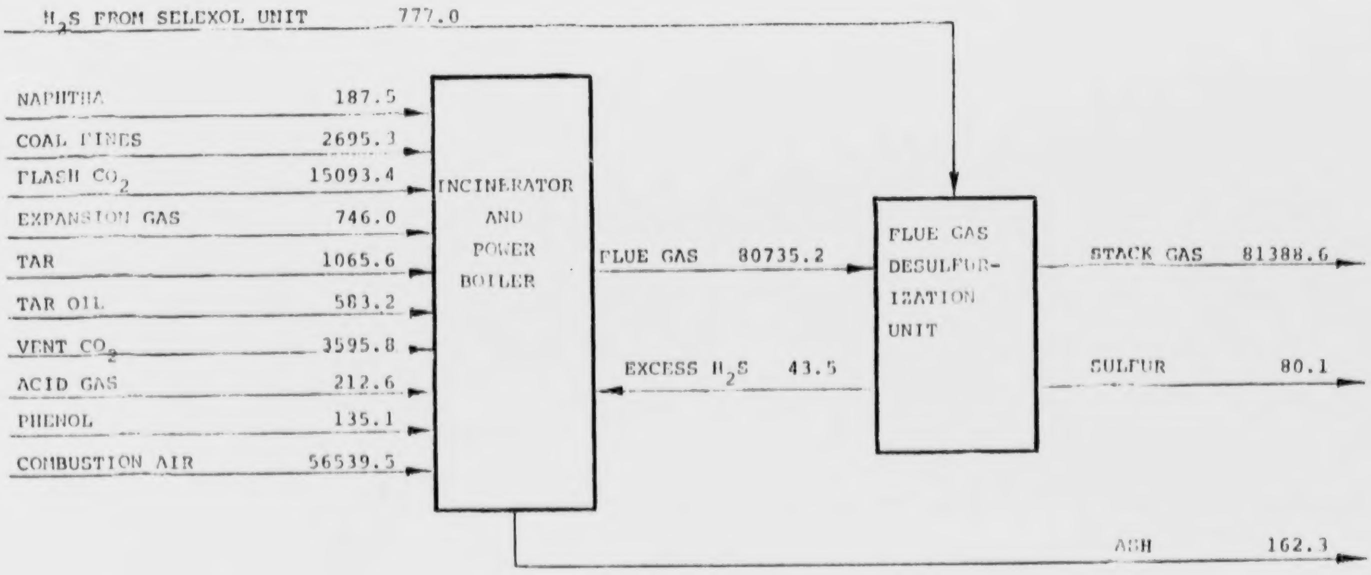
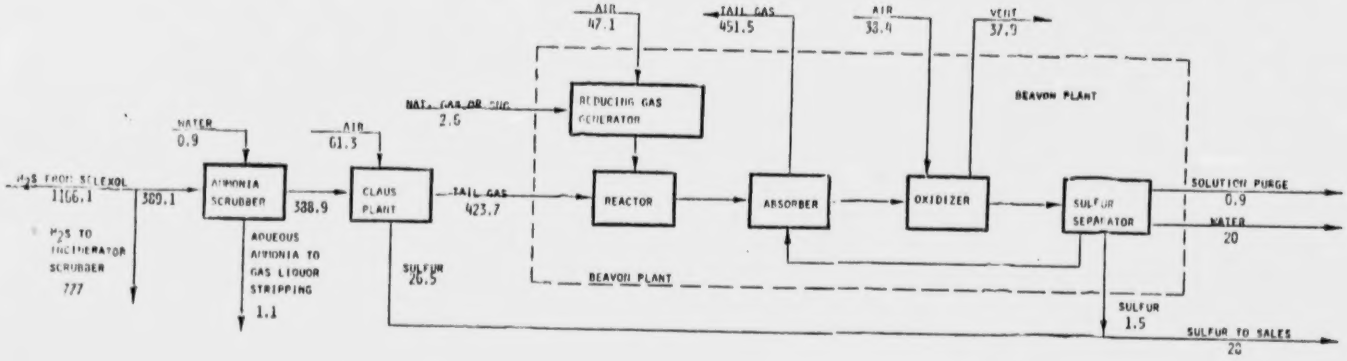

H2S FROM SELEXOL UNIT 777.0
NAPHTHA 187.5
COAL FINES 2695.3
FLASH CO2 15093.4
EXPANSION GAS 746.0
TAR 1065.6
TAR OIL 583.2
VENT CO2 3595.8
ACID GAS 212.6
PHENOL 135.1
COMBUSTION AIR 56539.5
INCINERATOR AND POWER BOILER
FLUE GAS 80735.2
EXCESS H2S 43.5
FLUE GAS DESULFURIZATION UNIT
STACK GAS 81388.6
SULFUR 80.1
ASH 162.3

Figure 9-8. Lurgi process: incinerator/power boiler and flue gas desulfurization.


H2S FROM SELEXOL
1166.1
389.1
H2S TO INCINERATOR SCRUBBER
777
WATER
0.9
AMMONIA SCRUBBER
338.9
AQUEOUS AMMONIA TO GAS LIQUOR STRIPPING
1.1
AIR
61.3
CLAUS PLANT
SULFUR
26.5
TAIL GAS
423.7
NAT. GAS OR SNG
2.6
AIR
47.1
REDUCING GAS GENERATOR
REACTOR
TAIL GAS
451.5
ABSORBER
AIR
33.4
OXIDIZER
VENT
37.9
BEAVON PLANT
SULFUR SEPARATOR
SOLUTION PURGE
0.9
WATER
20
SULFUR
1.5
SULFUR TO SALES
28
BEAVON PLANT

(FLOWS IN STPD)

Figure 9-9. Lurgi process: Claus-Beavon sulfur recovery.

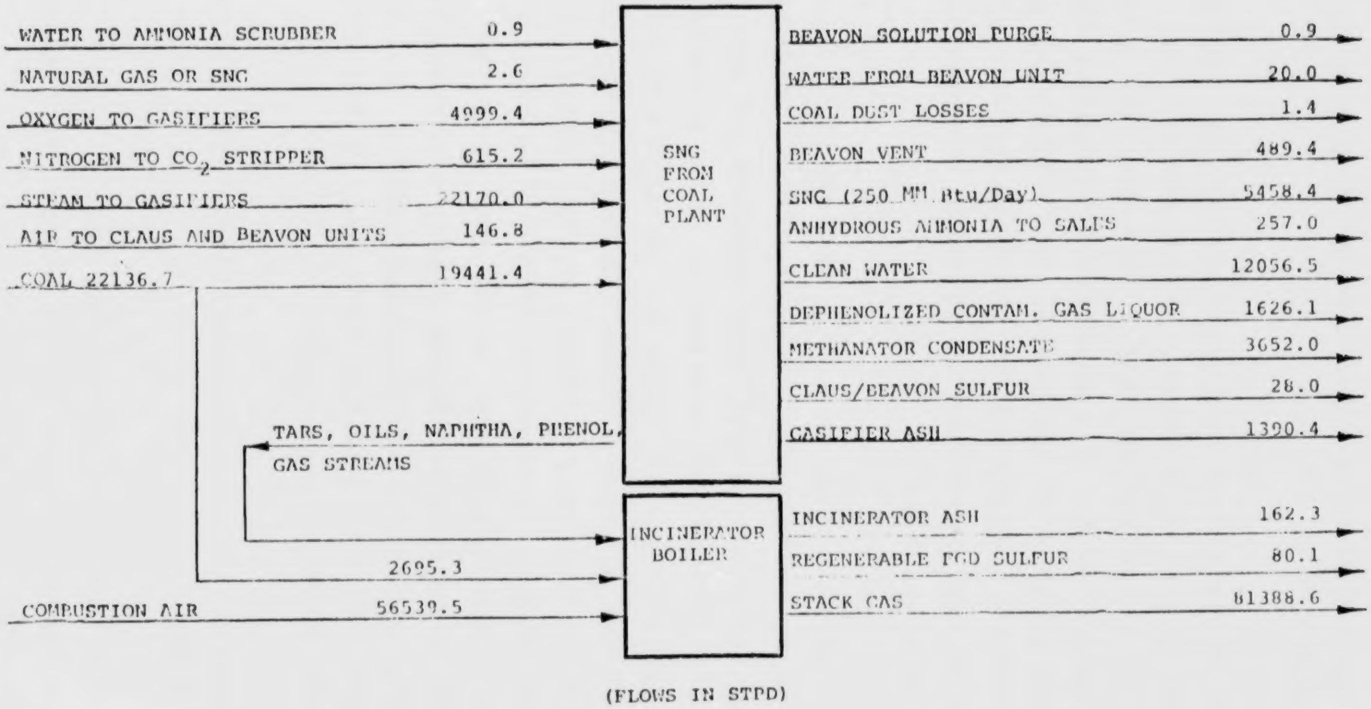

WATER TO AMMONIA SCRUBBER 0.9
NATURAL GAS OR SNG 2.6
OXYGEN TO GASIFIERS 4999.4
NITROGEN TO CO_2 STRIPPER 615.2
STEAM TO GASIFIERS 22170.0
AIR TO CLAUS AND BEAVON UNITS 146.8
COAL 22136.7
19441.4
TARS, OILS, NAPHTHA, PHENOL, GAS STREAMS
2695.3
COMBUSTION AIR 56539.5
SNG FROM COAL PLANT
INCINERATOR BOILER
BEAVON SOLUTION PURGE 0.9
WATER FROM BEAVON UNIT 20.0
COAL DUST LOSSES 1.4
BEAVON VENT 489.4
SNG (250 MM Btu/Day) 5458.4
ANHYDROUS AMMONIA TO SALES 257.0
CLEAN WATER 12056.5
DEPHENOLIZED CONTAM. GAS LIQUOR 1626.1
METHANATOR CONDENSATE 3652.0
CLAUS/BEAVON SULFUR 28.0
GASIFIER ASH 1390.4
INCINERATOR ASH 162.3
REGENERABLE FGD SULFUR 80.1
STACK GAS 61388.6
(FLOWS IN STPD)

Figure 9-10. Lurgi process: overall feed and product weight balance.

emissions of sulfur compounds into the atmosphere to meet environmental standards.

A similar series of calculations was carried out for the Bi-Gas high temperature gasification process, using the C. F. Braun study (1) as a basis, to establish sulfur balances for operation with high and low sulfur coal. The four sulfur balances are discussed later in this section of the report in "Integrated Schemes for Emissions Control". Figure 9-11 illustrates the gaseous emission streams from coal gasification to which control technology must be, or has been, applied.

The SRC II process was selected as being representative of the liquefaction processes, principally because of the availability of process data and information and also because of the collaboration with the liquid effluent studies in this project. The conceptual engineering design of Ralph M. Parsons (3) was used as a basis to typify the process streams and the gaseous emission streams. Figure 9-12 is the flowsheet of the SRC II process based on the Parsons design to convert 20,000 TPD of dry, high sulfur, eastern (Illinois, Indiana, Kentucky) coal fed to the dissolvers into SNG, liquefied petroleum gases, naphtha, and fuel oil. As in the Lurgi gasification flowsheet, sulfur recovery and sulfur emissions control systems are included in Figure 9-12, details of which will be discussed later.

Details of the sections of the liquefaction flow diagram are shown in the following:

Figure 9-13.	Coal slurrying and dissolving.
Figure 9-14.	Dissolver slurry fractionation, filtration, and naphtha hydrogenation.

Figure 9-15. Gas purification, drying, and separation.
Figure 9-16. Process gasifier and hydrogen production.
Figure 9-17. Fuel gas gasification, purification, and sulfur recovery.
Figure 9-18. Sour water stripping and ammonia recovery.
Figure 9-19. Sulfur recovery and tail gas treating.
Figure 9-20. Process furnaces and steam and power generation.

The overall feed and product weight balance for the plant is shown in Figure 9-21. The gaseous emission streams from coal liquefaction to which control technology must be, or has been, applied are shown in Figure 9-22. A sulfur balance considering the same high sulfur coal is discussed in "Integrated Schemes for Emissions Control" in this section of the report.

References

1. Detman, R., "Factored Estimates for Western Coal Commercial Concepts." October 1976. FE 2240-5. 294, 295, 296*
2. Sinor, J., "Evaluation of Background Data Relating to New Source Performance Standards for Lurgi Gasification." 1977. EPA 600/7-77-057. 552*
3. O'Hara, J., et al., "Oil/Gas Complex Conceptual Design/Economic Analysis: Oil and SNG Production." March 1977. FE-1775-B. 814*

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*Pullman Kellogg Reference File number.

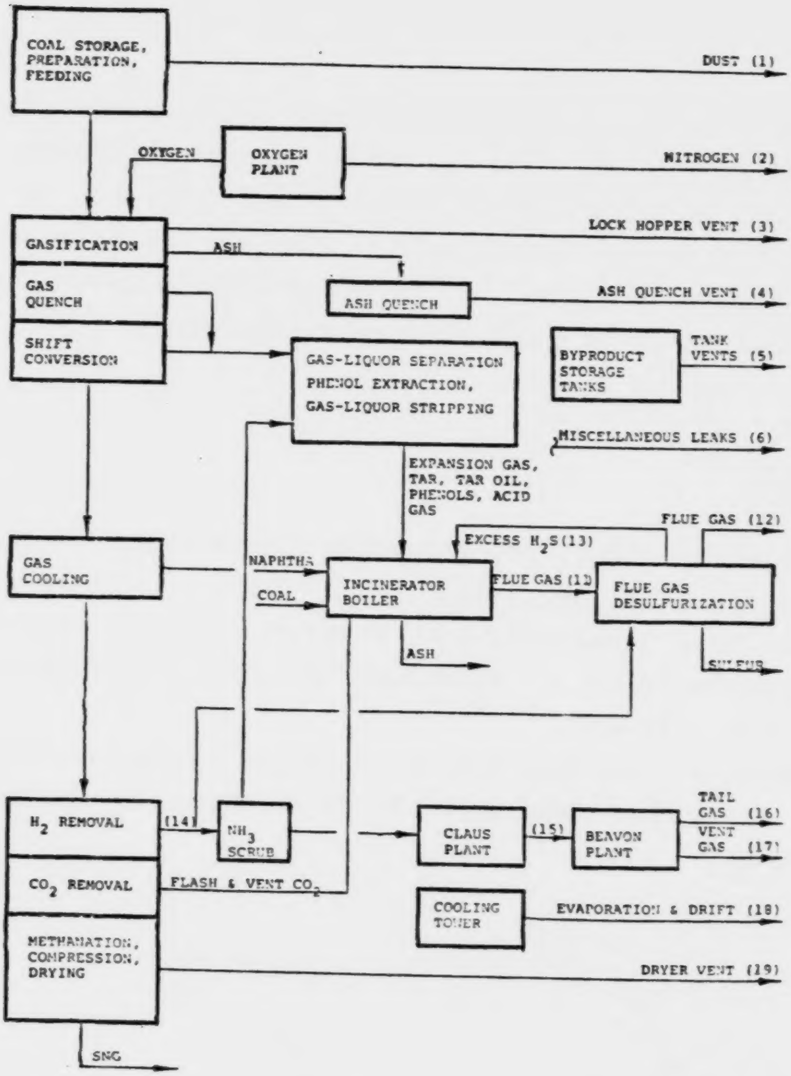
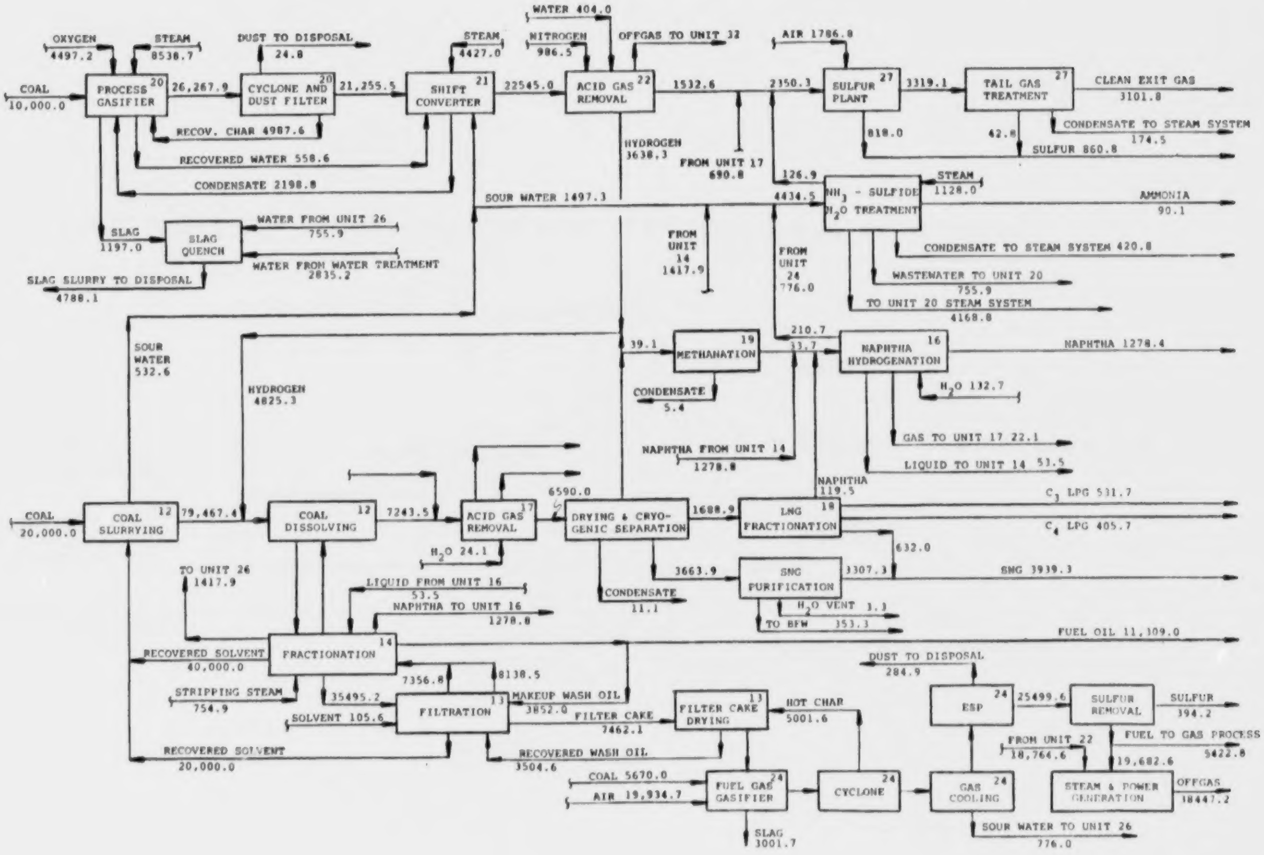

COAL STORAGE, PREPARATION, FEEDING
DUST (1)
OXYGEN
OXYGEN PLANT
NITROGEN (2)
GASIFICATION
LOCK HOPPER VENT (3)
ASH
GAS QUENCH
ASH QUENCH
ASH QUENCH VENT (4)
SHIFT CONVERSION
GAS-LIQUOR SEPARATION
PHENOL EXTRACTION,
GAS-LIQUOR STRIPPING
BYPRODUCT STORAGE TANKS
TANK VENTS (5)
MISCELLANEOUS LEAKS (6)
EXPANSION GAS, TAR, TAR OIL, PHENOLS, ACID GAS
FLUE GAS (12)
EXCESS H_2S(13)
GAS COOLING
NAPHTHA
COAL
INCINERATOR BOILER
FLUE GAS (11)
FLUE GAS DESULFURIZATION
ASH
SULFUR
H_2 REMOVAL
(14)
NH_3 SCRUB
CLAUS PLANT
(15)
BEAVON PLANT
TAIL GAS (16)
VENT GAS (17)
CO_2 REMOVAL
FLASH & VENT CO_2
COOLING TOWER
EVAPORATION & DRIFT (18)
METHANATION, COMPRESSION, DRYING
DRYER VENT (19)
SNG

Figure 9-11. Gaseous emission streams from coal gasification.



COAL 10,000.0
OXYGEN 4497.2
STEAM 8538.7
PROCESS GASIFIER 20
26,267.9
CYCLONE AND DUST FILTER 20
DUST TO DISPOSAL 24.8
21,255.5
RECOV. CHAR 4987.6
RECOVERED WATER 558.6
CONDENSATE 2198.8
SHIFT CONVERTER 21
STEAM 4427.0
22545.0
WATER 404.0
NITROGEN 986.5
ACID GAS REMOVAL 22
OFFGAS TO UNIT 32
HYDROGEN 3638.3
1532.6
FROM UNIT 17 690.8
2350.3
AIR 1786.8
SULFUR PLANT 27
3319.1
TAIL GAS TREATMENT 27
CLEAN EXIT GAS 3101.8
CONDENSATE TO STEAM SYSTEM 174.5
818.0
42.8
SULFUR 860.8
SLAG 1197.0
SLAG QUENCH
WATER FROM UNIT 26 755.9
WATER FROM WATER TREATMENT 2835.2
SLAG SLURRY TO DISPOSAL 4788.1
SOUR WATER 1497.3
FROM UNIT 14 1417.9
126.9
4434.5
NH₃ - SULFIDE H₂O TREATMENT
STEAM 1128.0
AMMONIA 90.1
CONDENSATE TO STEAM SYSTEM 420.8
WASTEWATER TO UNIT 20 755.9
TO UNIT 20 STEAM SYSTEM 4168.8
FROM UNIT 24 776.0
SOUR WATER 532.6
HYDROGEN 4825.3
39.1
METHANATION 19
CONDENSATE 5.4
210.7
33.7
NAPHTHA HYDROGENATION 16
NAPHTHA 1278.4
H₂O 132.7
GAS TO UNIT 17 22.1
LIQUID TO UNIT 14 53.5
NAPHTHA FROM UNIT 14 1278.8
NAPHTHA 119.5
COAL 20,000.0
COAL SLURRYING 12
79,467.4
COAL DISSOLVING 12
7243.5
ACID GAS REMOVAL 17
H₂O 24.1
6590.0
DRYING & CRYO-GENIC SEPARATION
CONDENSATE 11.1
1688.9
LNG FRACTIONATION 18
C₃ LPG 531.7
C₄ LPG 405.7
632.0
3663.9
SNG PURIFICATION
3307.3
SNG 3939.3
H₂O VENT 3.3
TO BFW 353.3
TO UNIT 26 1417.9
LIQUID FROM UNIT 16 53.5
NAPHTHA TO UNIT 16 1278.8
FRACTIONATION 14
FUEL OIL 11,309.0
RECOVERED SOLVENT 40,000.0
STRIPPING STEAM 754.9
35495.2
7356.8
8138.5
FILTRATION 13
SOLVENT 105.6
MAKEUP WASH OIL 3852.0
FILTER CAKE 7462.1
RECOVERED SOLVENT 20,000.0
RECOVERED WASH OIL 3504.6
FILTER CAKE DRYING 13
HOT CHAR 5001.6
COAL 5670.0
AIR 19,934.7
FUEL GAS GASIFIER 24
SLAG 3001.7
CYCLONE 24
GAS COOLING 24
SOUR WATER TO UNIT 26 776.0
ESP 24
DUST TO DISPOSAL 284.9
25499.6
SULFUR REMOVAL
SULFUR 394.2
FROM UNIT 22 18,764.6
19,682.6
FUEL TO GAS PROCESS 5422.8
STEAM & POWER GENERATION
OFFGAS 18447.2

Figure 9-12.- Coal liquefaction: SRC-II block flow diagram and material balance.

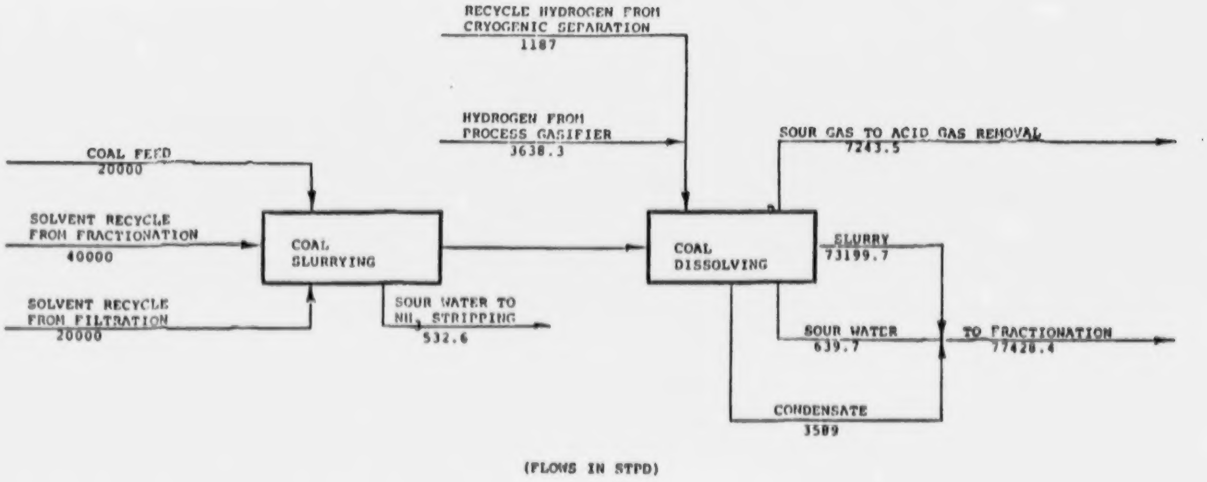

RECYCLE HYDROGEN FROM CRYOGENIC SEPARATION 1187
HYDROGEN FROM PROCESS GASIFIER 3638.3
COAL FEED 20000
SOLVENT RECYCLE FROM FRACTIONATION 40000
SOLVENT RECYCLE FROM FILTRATION 20000
COAL SLURRYING
SOUR WATER TO NH$_3$ STRIPPING 532.6
COAL DISSOLVING
SOUR GAS TO ACID GAS REMOVAL 7243.5
SLURRY 73199.7
SOUR WATER 639.7
TO FRACTIONATION 77428.4
CONDENSATE 3589
(FLOWS IN STPD)

Figure 9-13. Liquefaction: coal slurrying and dissolving.

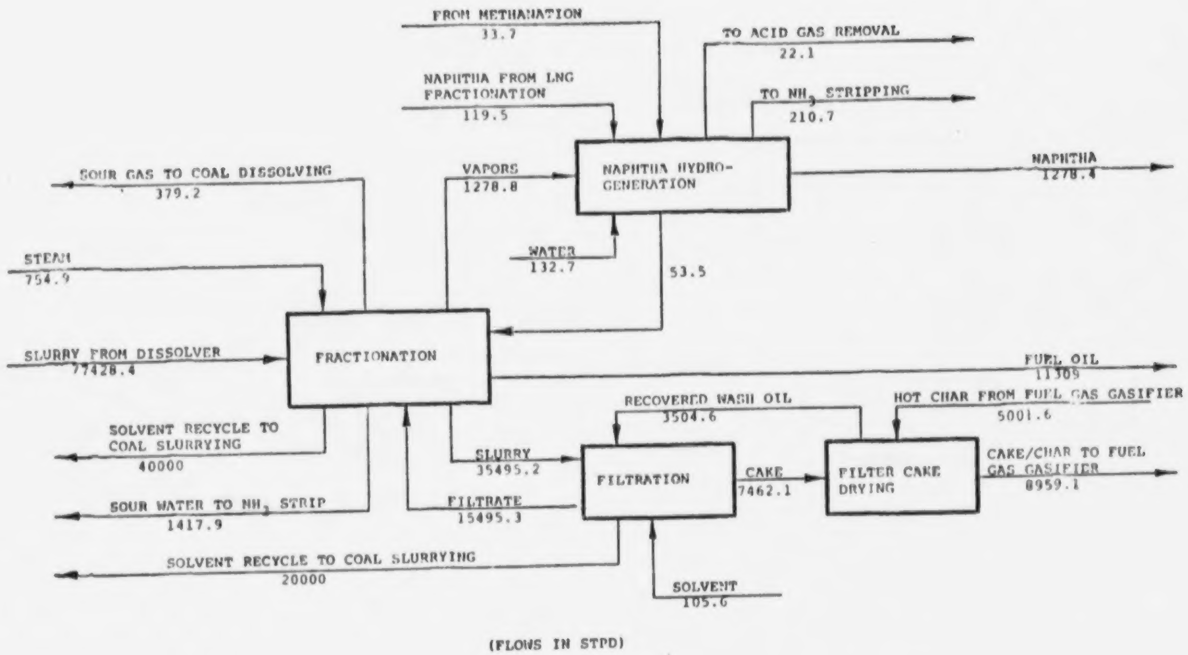
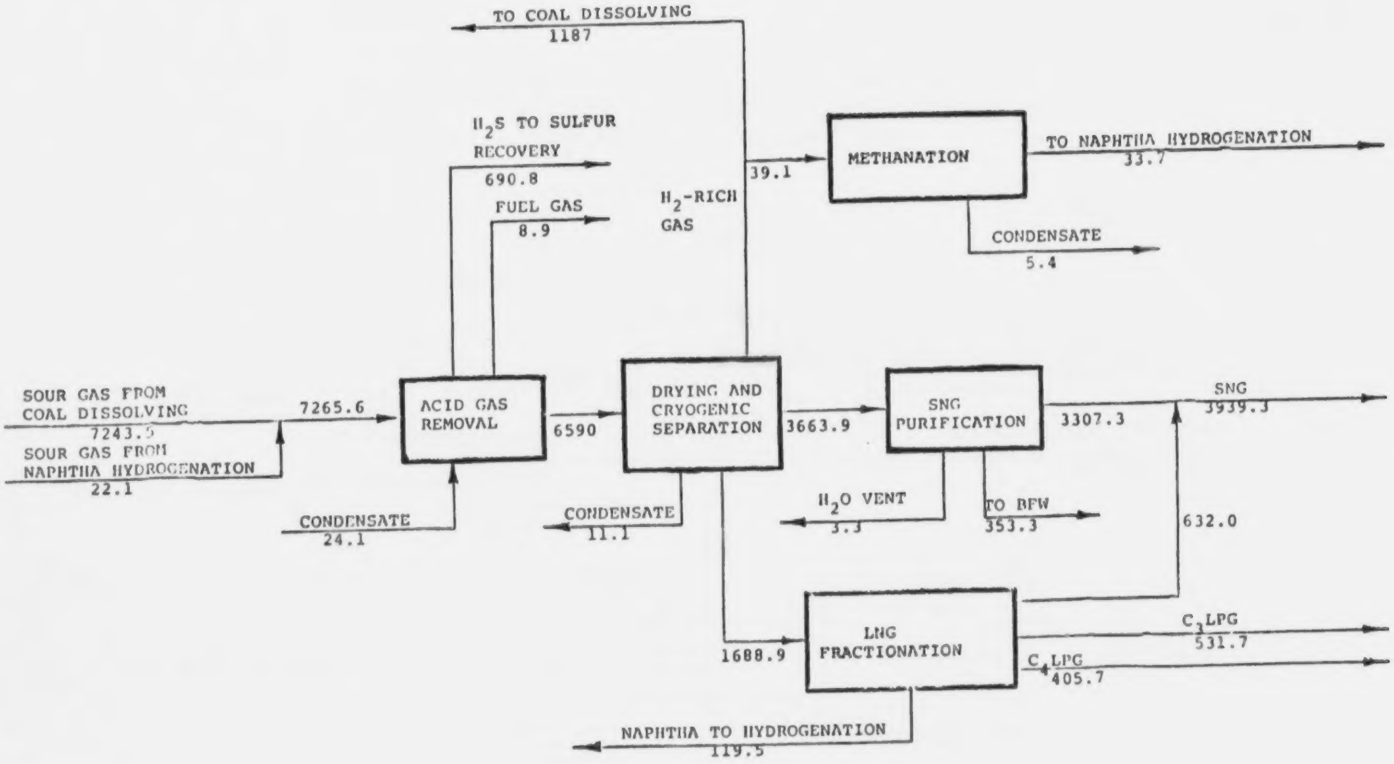
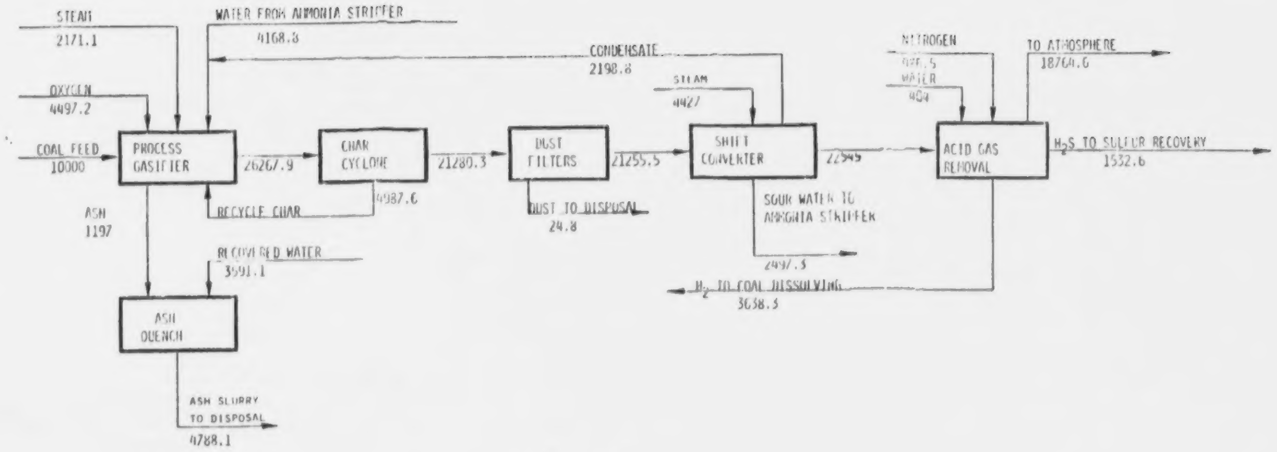

FROM METHANATION
33.7
NAPHTHA FROM LNG FRACTIONATION
119.5
TO ACID GAS REMOVAL
22.1
TO NH3 STRIPPING
210.7
SOUR GAS TO COAL DISSOLVING
379.2
VAPORS
1278.8
NAPHTHA HYDRO-GENERATION
NAPHTHA
1278.4
STEAM
754.9
WATER
132.7
53.5
SLURRY FROM DISSOLVER
77428.4
FRACTIONATION
FUEL OIL
11309
RECOVERED WASH OIL
3504.6
HOT CHAR FROM FUEL GAS GASIFIER
5001.6
SOLVENT RECYCLE TO COAL SLURRYING
40000
SLURRY
35495.2
FILTRATION
CAKE
7462.1
FILTER CAKE DRYING
CAKE/CHAR TO FUEL GAS GASIFIER
8959.1
SOUR WATER TO NH3 STRIP
1417.9
FILTRATE
15495.3
SOLVENT RECYCLE TO COAL SLURRYING
20000
SOLVENT
105.6
(FLOWS IN STPD)

Figure 9-14. Liquefaction: dissolver slurry fractionation, filtration, and naphtha hydrogenation.


TO COAL DISSOLVING
1187
H$_2$S TO SULFUR
RECOVERY
690.8
FUEL GAS
8.9
H$_2$-RICH
GAS
39.1
METHANATION
TO NAPHTHA HYDROGENATION
33.7
CONDENSATE
5.4
SOUR GAS FROM
COAL DISSOLVING
7243.5
SOUR GAS FROM
NAPHTHA HYDROGENATION
22.1
7265.6
ACID GAS
REMOVAL
CONDENSATE
24.1
6590
DRYING AND
CRYOGENIC
SEPARATION
CONDENSATE
11.1
3663.9
SNG
PURIFICATION
H$_2$O VENT
3.3
TO BFW
353.3
3307.3
SNG
3939.3
632.0
1688.9
LNG
FRACTIONATION
C$_3$LPG
531.7
C$_4$LPG
405.7
NAPHTHA TO HYDROGENATION
119.5

(FLOWS IN STPD)

Figure 9-15. Liquefaction: gas purification, drying and separation.


STEAM
2171.1
OXYGEN
4497.2
COAL FEED
10000
PROCESS GASIFIER
WATER FROM AMMONIA STRIPPER
4168.8
CONDENSATE
2198.8
26267.9
CHAR CYCLONE
4987.6
RECYCLE CHAR
ASH
1197
RECOVERED WATER
3591.1
ASH QUENCH
ASH SLURRY TO DISPOSAL
4788.1
21280.3
DUST FILTERS
DUST TO DISPOSAL
24.8
21255.5
STEAM
4427
SHIFT CONVERTER
SOUR WATER TO AMMONIA STRIPPER
2497.3
22545
NITROGEN
WATER
404
ACID GAS REMOVAL
TO ATMOSPHERE
18764.6
H_2S TO SULFUR RECOVERY
1532.6
H_2 TO COAL DISSOLVING
3638.3

(FLOWS IN STPH)

Figure 9-16. Liquefaction: process gasifier and hydrogen production.

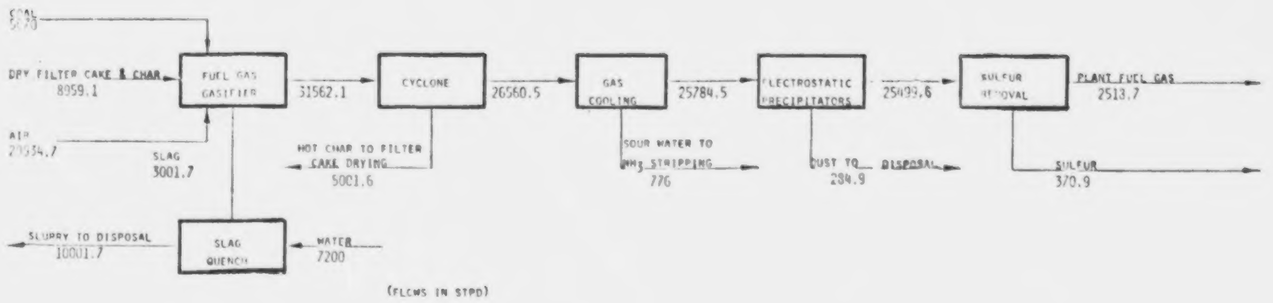
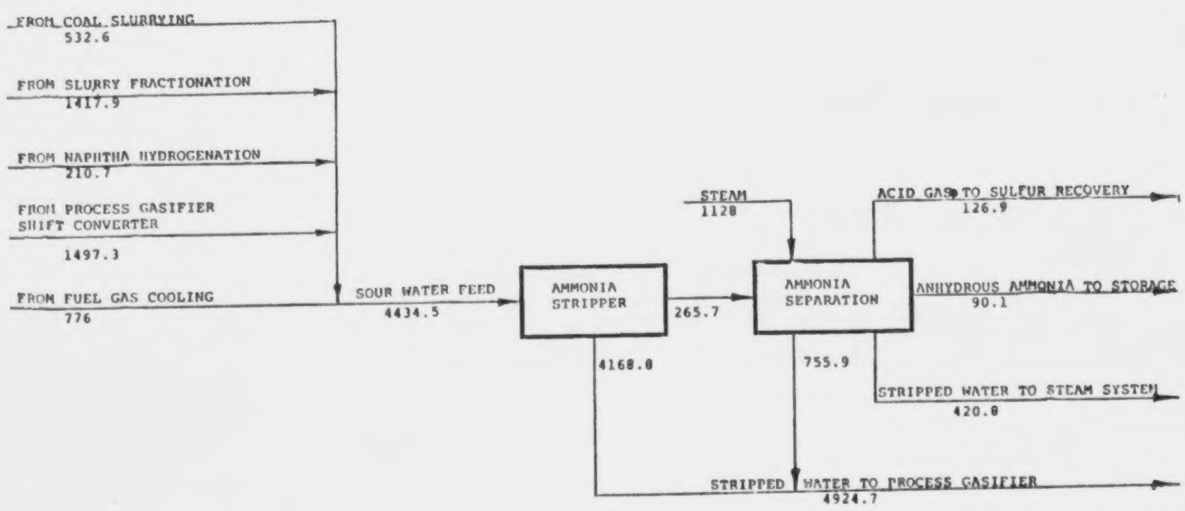
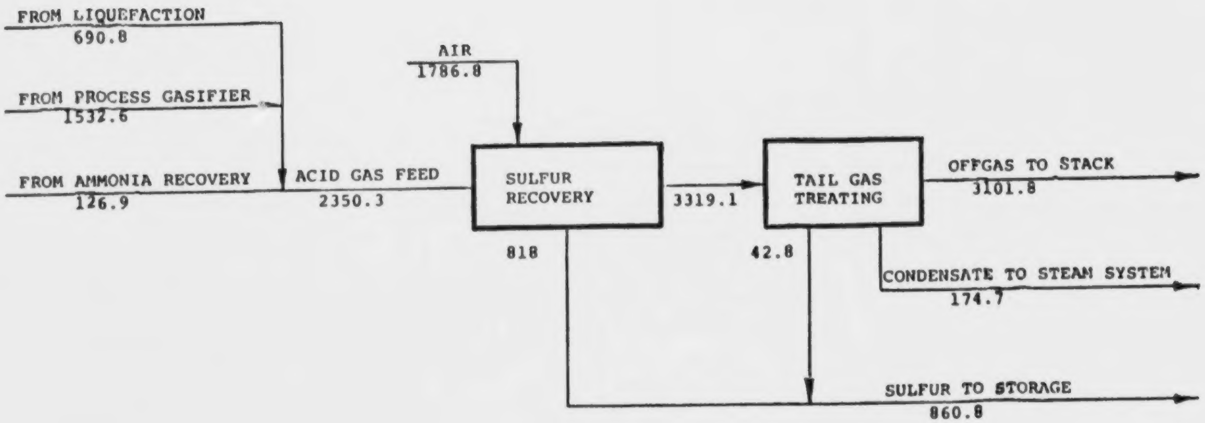

COAL
5670
DRY FILTER CAKE & CHAR
8959.1
AIR
20934.7
SLAG
3001.7
FUEL GAS GASIFIER
31562.1
CYCLONE
HOT CHAR TO FILTER CAKE DRYING
5001.6
26560.5
GAS COOLING
SOUR WATER TO NH3 STRIPPING
776
25784.5
ELECTROSTATIC PRECIPITATORS
DUST TO DISPOSAL
284.9
25499.6
SULFUR REMOVAL
PLANT FUEL GAS
2513.7
SULFUR
370.9
SLURRY TO DISPOSAL
10001.7
SLAG QUENCH
WATER
7200
(FLOWS IN STPD)

Figure 9-17. Liquefaction: fuel gas gasification, purification and sulfur recovery.


FROM COAL SLURRYING
532.6
FROM SLURRY FRACTIONATION
1417.9
FROM NAPHTHA HYDROGENATION
210.7
FROM PROCESS GASIFIER SHIFT CONVERTER
1497.3
FROM FUEL GAS COOLING
776
SOUR WATER FEED
4434.5
AMMONIA STRIPPER
265.7
4168.8
STEAM
1128
AMMONIA SEPARATION
755.9
ACID GAS TO SULFUR RECOVERY
126.9
ANHYDROUS AMMONIA TO STORAGE
90.1
STRIPPED WATER TO STEAM SYSTEM
420.8
STRIPPED WATER TO PROCESS GASIFIER
4924.7

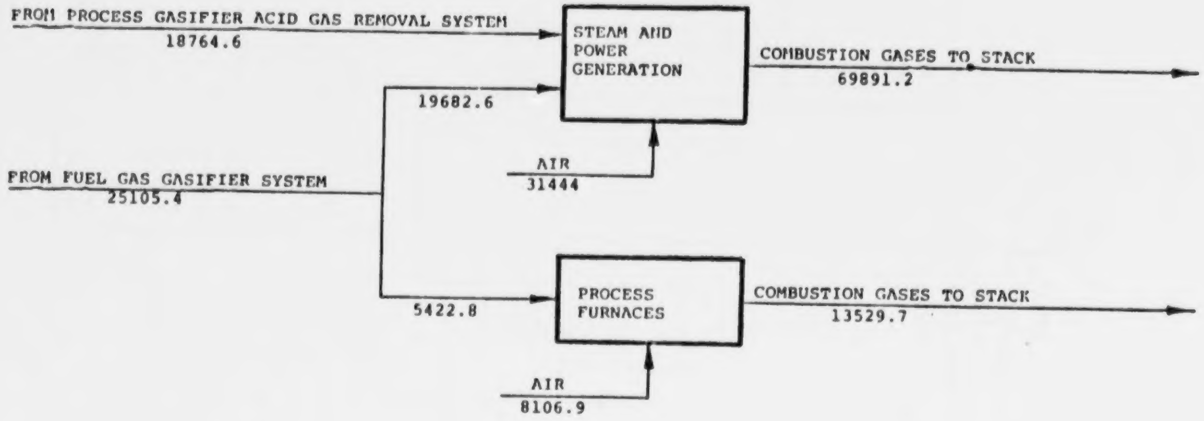
(FLOWS IN STPD)

Figure 9-18. Liquefaction: sour water stripping and ammonia recovery.


FROM LIQUEFACTION
690.8
FROM PROCESS GASIFIER
1532.6
FROM AMMONIA RECOVERY
126.9
ACID GAS FEED
2350.3
AIR
1786.8
SULFUR RECOVERY
818
3319.1
TAIL GAS TREATING
42.8
OFFGAS TO STACK
3101.8
CONDENSATE TO STEAM SYSTEM
174.7
SULFUR TO STORAGE
860.8

(FLOWS IN STPD)

Figure 9-19. Liquefaction: sulfur recovery and tail gas treating.


FROM PROCESS GASIFIER ACID GAS REMOVAL SYSTEM
18764.6
FROM FUEL GAS GASIFIER SYSTEM
25105.4
19682.6
5422.8
STEAM AND POWER GENERATION
PROCESS FURNACES
AIR
31444
AIR
8106.9
COMBUSTION GASES TO STACK
69891.2
COMBUSTION GASES TO STACK
13529.7

(FLOWS IN STPD)

Figure 9-20. Liquefaction: process furnaces and steam and power generation.

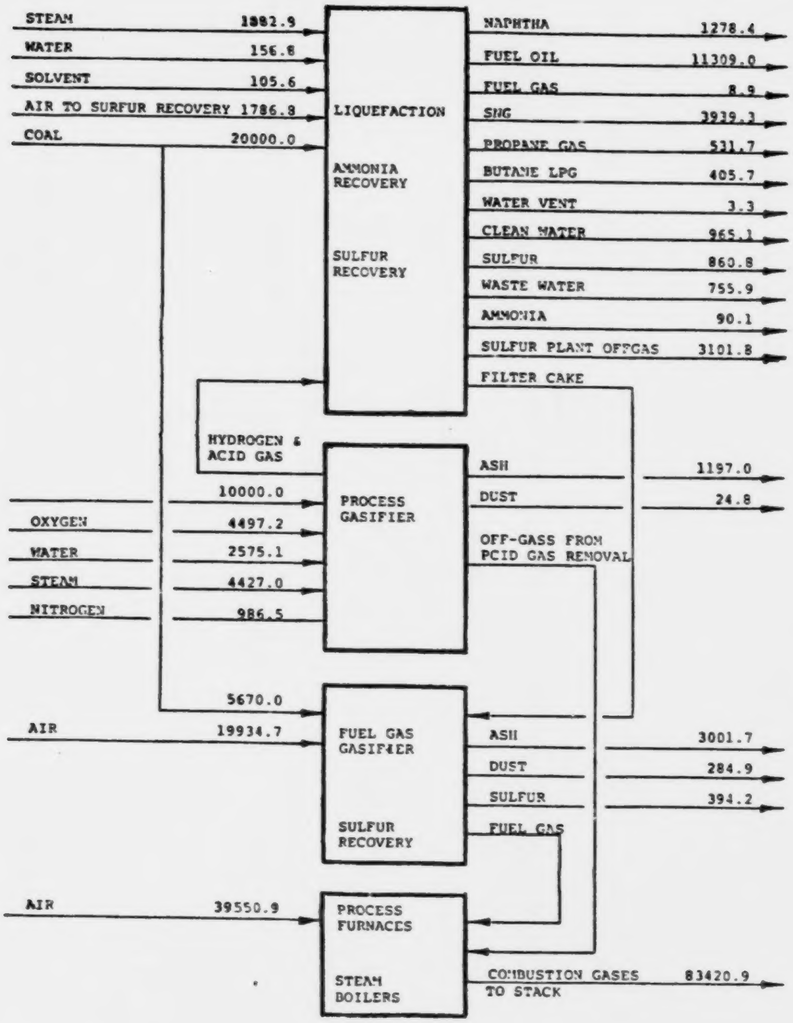

STEAM 1882.9
WATER 156.8
SOLVENT 105.6
AIR TO SURFUR RECOVERY 1786.8
COAL 20000.0
LIQUEFACTION
AMMONIA RECOVERY
SULFUR RECOVERY
NAPHTHA 1278.4
FUEL OIL 11309.0
FUEL GAS 8.9
SNG 3939.3
PROPANE GAS 531.7
BUTANE LPG 405.7
WATER VENT 3.3
CLEAN WATER 965.1
SULFUR 860.8
WASTE WATER 755.9
AMMONIA 90.1
SULFUR PLANT OFFGAS 3101.8
FILTER CAKE
HYDROGEN & ACID GAS
10000.0
OXYGEN 4497.2
WATER 2575.1
STEAM 4427.0
NITROGEN 986.5
PROCESS GASIFIER
ASH 1197.0
DUST 24.8
OFF-GASS FROM PCID GAS REMOVAL
5670.0
AIR 19934.7
FUEL GAS GASIFIER
SULFUR RECOVERY
ASH 3001.7
DUST 284.9
SULFUR 394.2
FUEL GAS
AIR 39550.9
PROCESS FURNACES
STEAM BOILERS
COMBUSTION GASES TO STACK 83420.9

Figure 9-21. Liquefaction: overall feed and product weight balance.

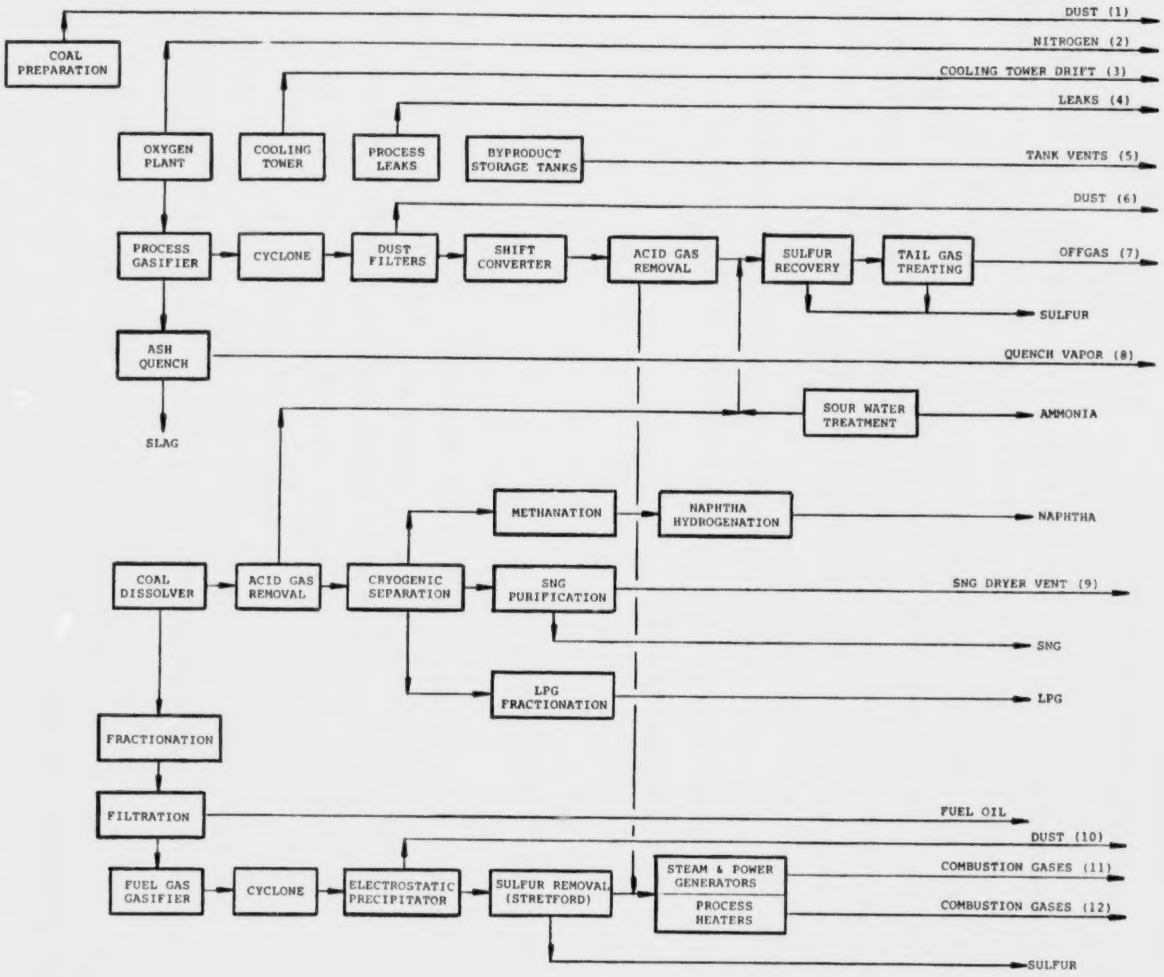

COAL PREPARATION
OXYGEN PLANT
COOLING TOWER
PROCESS LEAKS
BYPRODUCT STORAGE TANKS
PROCESS GASIFIER
CYCLONE
DUST FILTERS
SHIFT CONVERTER
ACID GAS REMOVAL
SULFUR RECOVERY
TAIL GAS TREATING
ASH QUENCH
SLAG
SOUR WATER TREATMENT
METHANATION
NAPHTHA HYDROGENATION
COAL DISSOLVER
ACID GAS REMOVAL
CRYOGENIC SEPARATION
SNG PURIFICATION
LPG FRACTIONATION
FRACTIONATION
FILTRATION
FUEL GAS GASIFIER
CYCLONE
ELECTROSTATIC PRECIPITATOR
SULFUR REMOVAL (STRETFORD)
STEAM & POWER GENERATORS
PROCESS HEATERS
DUST (1)
NITROGEN (2)
COOLING TOWER DRIFT (3)
LEAKS (4)
TANK VENTS (5)
DUST (6)
OFFGAS (7)
SULFUR
QUENCH VAPOR (8)
AMMONIA
NAPHTHA
SNG DRYER VENT (9)
SNG
LPG
FUEL OIL
DUST (10)
COMBUSTION GASES (11)
COMBUSTION GASES (12)
SULFUR

Figure 9-22.- Gaseous emission streams from coal liquefaction.

LITERATURE SURVEY AND DATA GATHERING

The project literature survey methods have been treated in Section 5 under the headings "Information Procurement, Storage, and Retrieval" and "Subjects Monitored."

Efforts in the collection of pertinent data and information on gaseous emissions from conversion processes closely paralleled the efforts of the water treating group in collection of liquid effluent data. Many of the same source reports and personal contacts produced information for both groups. Much additional information was gathered from Pullman Kellogg reports on control of gaseous emissions and designs for emission control processes, although great care was taken to avoid exposure of confidential information. These sources were supplemented by contacts with process and equipment vendors.

TARGET POLLUTANT RESIDUALS

As described in Section 6 of this report, the environmental standards group gathered into a separate report the federal, state, regional,and international regulations, both present and proposed, concerning permissible levels of contaminants in emissions to the atmosphere. A summary of the most stringent of these regulations was developed and was used as a standard for comparison of the efficiency of processes for emission control.

In TABLE 9-1 are shown the target pollutant residuals that were used as criteria in evaluation of emission control technology.

TABLE 9-1. TARGET POLLUTANT RESIDUALS: MOST STRINGENT AIR STANDARDS

Visible		
Any source	20% Opacity (No. 1 Ringelmann Chart)	All
Incinerator	10% Opacity (No. 0.5 Ringelmann Chart)	Mont.

Particulates	Heat Input MM Btu/hr	Standard, lb/MM Btu	
Coal fired	≥ 250	0.05	N.M.
	≤ 250	0.02 (≤ 2 microns dia.)	N.M.
	100	0.151	Colo.
	10	0.275	Colo.
	1	0.5	Colo.
Oil fired	≥ 114	0.005	N.M.
	≤ 114	0.10	Ill.
Gas fired	$> 2,500$	0.10	Tex.
Combined fuels	Any Size	$E=S_sH_3 + 0.10\ H_\ell$ where E=Allowable particulate emission, lb/MM Btu S_s=Solid fuel particulate emission standard, lb/MM Btu H_3=Actual heat input from solid fuel, MM Btu/hr H_ℓ=Actual heat input from liquid fuel, MM Btu/hr.	Ill.

Fugitive Dust		
	None visible outside property line	Kan.

Nitrogen Oxides	Heat Input, MM Btu/hr	Standard, lb NO_2MM Btu	
Coal fired	≥ 250	0.45	N.M.
Oil fired	≥ 250	0.30	Most
Gas fired	≥ 250	0.20	Most
Combined fuels	Any Size	$E=(0.2X+0.3Y+0.7Z)/(X+Y+Z)$ where E=lb NO /MM Btu X,Y,Z=% of total heat input from gas, liquid and solid fuels, respectively.	Colo.

TABLE 9-1. TARGET POLLUTANT RESIDUALS: MOST STRINGENT AIR STANDARDS (CONT)

Hydrocarbon Vapors

Organic materials	15 lb/day, 3 lb/hr*	Colo.
PCR (photochemically reactive)	40 lb/day, 8 lb/hr*	Colo.
Non-PCR	3,000 lb/day, 450 lb/hr*	Colo.

*Unless reduced by 85%

Other Chemicals

Fluoride	6 ppb(v) as HF, 3 hr. ave.	Tex.
Beryllium	0.01 μg/m^3, 24 hr. ave.	Tex.
Nitric acid mist and/or vapor	70 mg/500 Nm^3	W. Va.
Hydrochloric acid mist and/or vapor	210 mg/500 Nm^3	W. Va.
Mercury	7.05 lb/24 hr. from incineration of wastewater treatment sludges	Fed.

Sulfur Dioxide (Fossil fuel fired steam generators)

	Heat Input, MM Btu/hr	Standard, lb/MM Btu	
Coal fired	≤ 250	1.2	Okla.
	≥ 250	0.2	Wyo.
Oil fired, ASTM grades 4, 5, 6	≤ 115	1.0	Ill.
	≥ 115	0.34	N.M.
Oil fired, ASTM grades 1, 2	≤ 250	0.3	Ill.
	≥ 250	0.3	Ill.
Gas fired		0.13	Okla.
		50 grains H_2S/100 SCF exit gas	Ill.
Combined Fuels	Any Size	$E=S_sH_s+0.3H_d+S_rH_r$ where E=lb SO_2/hr S_s=Solid fuel emission standard, lb/MM Btu H_s=Actual heat input from solid fuel, MM Btu/hr H_d=Actual heat input from distillate oil fuel (ASTM grades 1, 2), MM Btu/hr S_r=Emission standard for residual oil fuel (ASTM grades 4, 5, 6), lb/MM Btu H_r=Actual heat input from residual oil fuel, MM Btu/hr	Ill.

TABLE 9-1. TARGET POLLUTANT RESIDUALS: MOST STRINGENT AIR STANDARDS (CONT)

Carbon Monoxide (Corrected to 50% excess air)		
Fuel combustion	200 ppm(v) CO	Ill.
Incinerators	500 ppm(v) CO	Ill.
Petroleum processes	350 ppm(v) CO	Ill.
Petroleum processes, FCCRU*	350 ppm(v) CO	Ill.
Odors	Prevented beyond property line	All
Waste Gas Disposal	Flares to be smokeless	All
Gasification Plants		
SO_2, Gas burning boilers	0.16 lb SO_2/MM Btu(LHV)	N.M.
Sulfur	0.008 lb S/MM Btu(HHV)	N.M.
H_2S, COS, CS_2	100 ppm(v) in effluent gas	N.M.
H_2S component	10 ppm(v) in effluent gas (max.)	N.M.
Ammonia	25 ppm(v) in effluent gas	N.M.
HCl	5 ppm(v) in effluent gas	N.M.
HCN	10 ppm(v) in effluent gas	N.M.
Sulfur Recovery Plants		
Sulfur dioxide	0.01 lb SO_2/lb sulfur processed	O., Ala.
	100 lb SO_2/hr (max.)	Okla.
Hydrogen sulfide	150 ppm(v) in effluent gas	Ala.
Petroleum Processing Facilities		
Hydrogen sulfide	10 ppm(v) in effluent gas	N.M.
Fuel gas burning equip.	0.1 grain/500 SCF in fuel gas	N.M.
Mercaptans	0.25 lb/hr total mercaptans	N.M.
Ammonia	25 ppm(v) in effluent gas	N.M.
H_2SO_4/SO_3		
Sulfuric acid plants	35 mg/Nm^3 of effluent gas	Mo.
Other than sulfuric acid plants	<1,300 TPY H_2SO_4 usage: 0.1 lb/hr	Ill.
	$\geq$1,300 TPY H_2SO_4 usage 0.5 lb/T H_2SO_4 used	Ill.
General Processing Facilities		
Sulfur dioxide	500 ppm(v)	Colo. Ohio Pa.

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*FCCRU= Fluid catalytic cracking regeneration unit

TREATMENT OF GASEOUS EMISSIONS AS SUGGESTED BY OTHERS

In the discussion that follows, the treatment methods proposed for the emission streams shown in Figures 9-11, for gasification, and 9-22, for liquefaction, in conceptual designs and process analyses are compared with the target pollutant residuals.

Dust (Streams 1)

Cameron Engineers (3, p.22) estimated particulate emissions to be 0.05 lb/ton of coal for crushing, screening, and conveying, plus 0.025 to 0.04 lb/ton of coal for storage and reclaiming, based on information obtained from the Wyoming Coal Gas Company. In addition to use of water sprays with a wetting agent at dump hoppers, transfer points, screens and crushers, and use of dust collectors in the screening plant, Cameron suggested that screening and coal fines cleaning operations be enclosed and vented to wet scrubbers or baghouses, that conveyors be covered, that free fall of coal onto the storage pile be minimized and that spontaneous combustion in storage piles be prevented by avoiding segregation of fines and compaction. These or similar methods are recommended by Exxon Research and Engineering Company (1), Battelle (2), by Pullman Kellogg in a client study (4) and by Parsons (5).

A critique of these methods is included in Section 10 of this report.

Nitrogen (Streams 2)

Cameron and all others proposed that waste nitrogen from the oxygen plant be vented to atmosphere through a 150 to 300 foot stack to disperse the nitrogen. It is felt that this method is adequate to avoid dangerous concentrations of nitrogen at ground level, since its lower molecular weight will compensate for the low exhaust temperature of about 7°F to yield a resultant density lower than that of ambient air.

Lock Hopper Vent Gas (Stream 3, Gasification)

Cameron examined several alternate schemes, of which the most desirable were (a) use of crude gas to pressurize the hoppers and a high pressure compressor to recycle most of the gas and (b) use of crude gas to pressurize with a low pressure recycle compressor. Residual vent gases were to be incinerated to recover heating value. The flue gases were to be scrubbed if necessary for SO_2 control.

Battelle, Exxon,and Kellogg proposed to use crude gas for pressurizing with a recycle compressor. Residual gases were to be vented to the atmosphere.

The venting of residual lock hopper gas to atmosphere will violate environmental standards. The most practical, economical, and environmentally satisfactory method is displacement of the residual gases with an inert gas stream, either CO_2 or nitrogen, to the incinerator before the lock hopper is opened.

Ash Quench Vent (Streams 4)

Cameron Engineers proposed to quench the high pressure ash lock vent gases in a direct contact water condenser, since these gases are mainly steam. Quenching of the ash evolves a large quantity of steam that carries ash and clinkers. Camerson proposed to separate the solids in a wet cyclone and then condense the steam with collection of fine ash particles in a direct contact condenser.

Alternately, Cameron noted that if the ash lock chamber is not repressurized before the valve to the gasifier is opened, gas from the gasifier could flow into the hopper and be emitted with the other noncondensables from the condenser.

Battelle advocated venting the condenser exit gases to atmosphere through a 100-to 200-foot stack. Kellogg proposed a final water spray into the vent gases to remove any remaining dust.

Parsons returned the vapors from the process gasifier quench to the gasifier. No disposal method is described for vapors from the fuel gasifier quench.

The control method selected after study of these proposals uses cyclones to collect cinders, a direct contact condenser to condense steam, and incineration of the noncondensable gases in order to oxidize any organic materials prior to discharge to the atmosphere.

Tank Vents and Miscellaneous Leaks (Streams 5 and 6)

The emissions from tank breathing, leaks, spills, and venting of tanks during filling can be any of the byproducts and chemicals that must be stored. Cameron estimated the more important emissions, based on API design, to be about 12.5 lb/hr, consisting of crude phenol 12 percent, tar oil 21 percent, naphtha 17 percent, ammonia 12 percent, product gases 25 percent and methanol 13 percent.

Cameron's proposed methods for controlling tank ventings included a refrigerated vent condenser, scrubbing with a low vapor pressure solvent, incineration, and adsorption. Exxon presented similar tank vent figures but did not propose control methods. Battelle and Kellogg did not mention quantities or means of control.

Parsons proposed refrigerated LPG storage with evaporation recovery.

Evaluation of these proposals showed that the Cameron combination of methods appeared to offer the best assurance that environmental standards will be met. It is felt that the possibility of hydrocarbon losses could be reduced by using floating roof storage tanks equipped with secondary (wiper) seals.

CO_2 from Acid Gas Removal (Streams 7 and 8)

The CO_2 streams from acid gas removal in gasification are routed to vent stacks for dispersion into the atmosphere in the Cameron, Exxon, and Battelle reports. Pullman Kellogg's report recommended routing the CO_2-rich gases to the utility boiler for incineration prior to venting to the atmosphere. Parsons advocates incineration and venting to atmosphere.

Because the CO_2-rich gases from both the Selexol and Rectisol units contain substantial quantities of H_2S, CO, CH_4, C_2H_4, and C_2H_6, venting directly to the atmosphere is environmentally unacceptable. Therefore, incineration prior to venting was selected as being the most practical and acceptable means of control.

Expansion Gas and Acid Gas (Streams 9 and 10)

Exxon advocates sending the expansion gases from gas-liquor separation and acid gases from gas-liquor stripping to incineration with flue gas desulfurization (FGD) prior to venting to the atmosphere. These gases are not mentioned in the Battelle and Pullman Kellogg reports. Expansion and acid gases in liquefaction were routed by Parsons to the Claus sulfur recovery plant.

Incineration followed by efficient FGD appears to be practical and adequate for environmental acceptability.

Incinerator/Boiler Flue Gas (Stream 11) and Flue Gas Desulfurization (Stream 12)

The Exxon and Battelle reports proposed that the combined flue gases from the incinerator and the superheater be sent to an FGD system prior to being vented to the atmosphere, with Battelle recommending a wet limestone scrubber for FGD. Pullman Kellogg proposed venting direct to the atmosphere, with the precaution that the boiler feed gases should be thoroughly mixed to avoid bypassing to zones of lower temperature with consequent incomplete incineration. Parsons advocated desulfurization of the fuel gas prior to combustion and then venting the flue gases to atmosphere.

Study of the compositions of the streams to the incinerator/-boiler led to the conclusion that either FGD or desulfurization of the feed gases before incineration would be adequate to meet most environmental standards, with final choice being made from economic studies.

Particulates must be removed from the flue gases by cyclones and electrostatic precipitators. NO_x emissions may be controlled by boiler modifications or by flue gas treatment.

Sulfur Recovery Tail Gas (Stream 13)

Battelle proposed to recover sulfur in a Claus plant, incinerate the tail gas, remove SO_2 in a wet limestone scrubber, and then vent the gases to atmosphere through 150- to 300-foot stacks. Exxon proposed to recover sulfur with the Stretford process, incinerate the tail gas in the superheater boiler and treat in an

FGD system before venting to atmosphere. Pullman Kellogg advocated recovery of sulfur in a Stretford plant, incineration of the offgases in the utility boiler, and venting to atmosphere. Cameron proposed to recover sulfur in a Stretford unit, incinerate and vent the Stretford tail gas, and vent the lean absorber and oxidizer offgases directly to atmosphere. Parsons advocated sulfur recovery in a Claus plant with tail gas treatment in a Beavon plant, with Beavon tail gas vented to atmosphere.

Cameron proposed an alternate scheme in which the rich H_2S stream to sulfur recovery would be treated in a Claus plant and the lean H_2S stream would be treated in a Stretford plant. Tail gases from both processes would be incinerated and scrubbed before release. In a second alternate, Cameron proposed to treat only the tail gas from the Claus plant, then the offgases from tail gas treatment and tail gas from the Stretford plant would be combined, incinerated, and vented.

Evaluation of these various proposed processing schemes led to the conclusions that:

- Present environmental standards for many states would probably be met by incineration and venting to the atomosphere.
- Present most stringent environmental standards would only be met by use of the Beavon, or similar process, or by tail gas incineration and FGD.

Evaluation of the entire sulfur handling problem led to the conclusions that:

- Recovery of sulfur in salable form would probably be more economically attractive than FGD with a throwaway process.
- Claus plant tail gas can be economically treated to meet most stringent environmental standards in a Beavon plant, with additional sulfur recovery.

- Because of the availability of an H_2S-rich gas stream in the coal conversion plants, incinerator/boiler flue gases can be economically treated to environmentally acceptable levels with the citrate process, with additional recovery of sulfur.

Accordingly, the Claus/Beavon and the citrate processes were selected for further study in integrated process schemes.

SNG Dryer Vent (Stream 14)

Exxon proposed to cool the gas leaving methanation, separate the condensate, remove water and CO_2 from the gas stream in the acid gas removal system,and then compress the dried gas to pipeline pressure. Cameron proposed to remove water from the SNG via glycol dehydration and then recover liquid water from the overhead of the glycol regenerator column. Parsons advocates glycol dehydration but vents the water vapor to atmosphere.

The SNG dryer vent gas consists of water vapor with about 8 mole percent methane and, because of the small quantity involved, can be vented through stacks without creating environmental problems.

Cooling Tower Evaporation and Drift (Stream 15)

The reports consider only the water content of the evaporation and drift from the cooling tower system, except where Cameron notes traces of ammonia and non-methane hydrocarbons.

Since the cooling tower in these reports is a receiver of blowdowns and wastewaters that may contain volatile contaminants in appreciable quantities, and since the cooling tower water may contain substantial quantities of nonvolatile materials that may be noxious or hazardous or cause nuisances, it is concluded that possible problems may have been dismissed too lightly. Accordingly, in studies of integrated processing schemes for gaseous emissions the means of reducing cooling tower drift were investigated and for liquid effluents the means of rendering cooling water constituents nonvolatile and nontoxic were developed.

References

1. Shaw, H., and Magee, E., "Evaluation of Pollution Control in Fossil Fuel Conversion Processes; Gasification; Section 1. Lurgi Process." EPA 650-2-74-009. 35*

2. Battelle, "Detailed Environmental Analysis Concerning a Proposed Coal Gasification Plant for Transwestern Coal Gasification Co. Pacific." Feb 1, 1973. 476*

3. Sinor, J., "Evaluation of Background Data Relating to New Source Performance Standards for Lurgi Gasification." June, 1977. 552*

4. Pullman Kellogg, "Engineering Evaluation of a Process to Produce 250 Billion Btu/Day of Pipeline Quality Gas From Illinois and Wyoming Coal for Panhandle Eastern Pipeline Company and Peabody Coal Company." June 20, 1972.

5. O'Hara, J. B., and Hervey, G. H., et al., "Oil/Gas Complex Conceptual Design/Economic Analysis: Oil and SNG Production." R and D Report No. 114, Interim Report No. 4. March 1977. 814*

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*Pullman Kellogg Reference File number

COMMERCIAL EMISSION CONTROL METHODS

By definition, commercial emission control methods are those processes and equipment that have been operated in full scale applications. The processes are offered by licensors who are, in most cases, the developers and operators of the processes. The equipment is offered by vendors who are able to demonstrate commercial operations.

The emission control methods that are considered to be most important for application to coal conversion processes, and that have been investigated, are:

- Processes and techniques for control of nitrogen oxides (NO_x)
- Processes and techniques for control of sulfur dioxide
- Processes for control of hydrogen sulfide
- Techniques for control of particulates
- Control of cooling tower drift
- Other control techniques applied to hydrocarbons, lock hopper vent gas, ammonia, ash quench vent gas, and miscellaneous leaks.

The emission control methods are presented in a general format that includes discussion of:

- Process description
- Capability, efficiency, and limitations
- Case histories
- Wastes produced
- Cost data (As found in the literature. Later in this section of the report these costs will be updated to the same year and on the same basis for comparison purposes.)
- Possible problems
- Possible improvements
- References

Processes and Techniques for Control of Nitrogen Oxides

Nitrogen oxides (NO_x) are formed during the combustion of coal, oil, and gas with air. The NO_x is comprised of about 95 percent nitric oxide, NO, and about 5 percent nitrogen dioxide, NO_2 (1,p.20).[*] Nitric oxide is colorless while nitrogen dioxide has a reddish-brown color. Both are toxic and both undergo photochemical reactions with hydrocarbons in the atmosphere to produce hignly active free radicals which create aerosols and toxicants that typify photochemical smog.

NO_x is formed by two mechanisms:

- high temperature fixation of oxygen and nitrogen present in the combustion air (thermal NO_x)
- reaction of oxygen with nitrogen contained in the fuel (fuel NO_x)

In either case, NO is the primary product formed because the residence time in most stationary combustion units is too short for a significant amount of NO to be oxidized to NO_2.

Formation of thermal NO_x is primarily dependent on temperature, being greater at higher temperatures. Its formation is also affected by oxygen availability. Formation of fuel NO_x, on the other hand, is relatively independent of combustion temperature but dependent on the amount of nitrogen in the fuel and the availability of free oxygen outside the combustion zone. NO_x

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[*] Other oxides of nitrogen formed to a minor degree are NO_3, N_2O, N_2O_3, N_2O_4, and N_2O_5 (2,p.7)

will be formed by both mechanisms during the combustion of fossil fuels although it is difficult to predict the exact contribution of each. Less than 30 percent of the nitrogen contained in coal is normally converted to NO_x (2,p.9); however, about 40 to 70 percent of that contained in oil will be converted to NO_x.

As shown in TABLE 9-2 and Figure 9-23, the percentage of liquid fuel nitrogen converted to NO_x is a function of the nitrogen content of the fuel. Fuels with a relatively high nitrogen content, 0.4 to 1.0 percent, will exhibit a lower percentage conversion to NO_x than those with the relatively low nitrogen content of 0.1 to 0.2 percent.

Some NO undergoes decomposition via the following reaction in the hot zones of the boiler:

$$2NO_{(g)} \rightarrow N_{2(g)} + O_{2(g)}$$

However, once rapid cooling of the gases begins, no further decomposition occurs. NO is continuously oxidized to NO_2 as follows:

$$NO_{(g)} + 1/2\ O_{2(g)} \rightarrow NO_{2(g)}$$

This reaction is slow and the time available is too short for equilibrium to be reached. Thus, as mentioned above, most of the NO_x leaving the boiler will be in the form of NO.

Currently, the most stringent standard for NO_x emissions from fuel combustion is that of New Mexico:

	NO_x, lb/MM Btu	NO_x, ppm (v)
Coal	0.45	338
Oil	0.30	225
Gas	0.20	150

TABLE 9-2. CONVERSION OF FUEL NITROGEN TO NO_x*

Fuel Type	Nitrogen Content, Wt.%	Conversion to NO_x, %
#2 Oil	0.006-0.02	100**
#5 Oil	0.10	56-60
#5 Oil	0.13	70
#5 Oil	0.20	41
#5 Oil	0.28	44
#6 Oil	0.27	52
#6 Oil	0.44	43-51
#6 Oil	1.50	45

* From Item 3, p. 96 in reference list

**Fuel nitrogen content too low to determine realistic values. Near 100%.

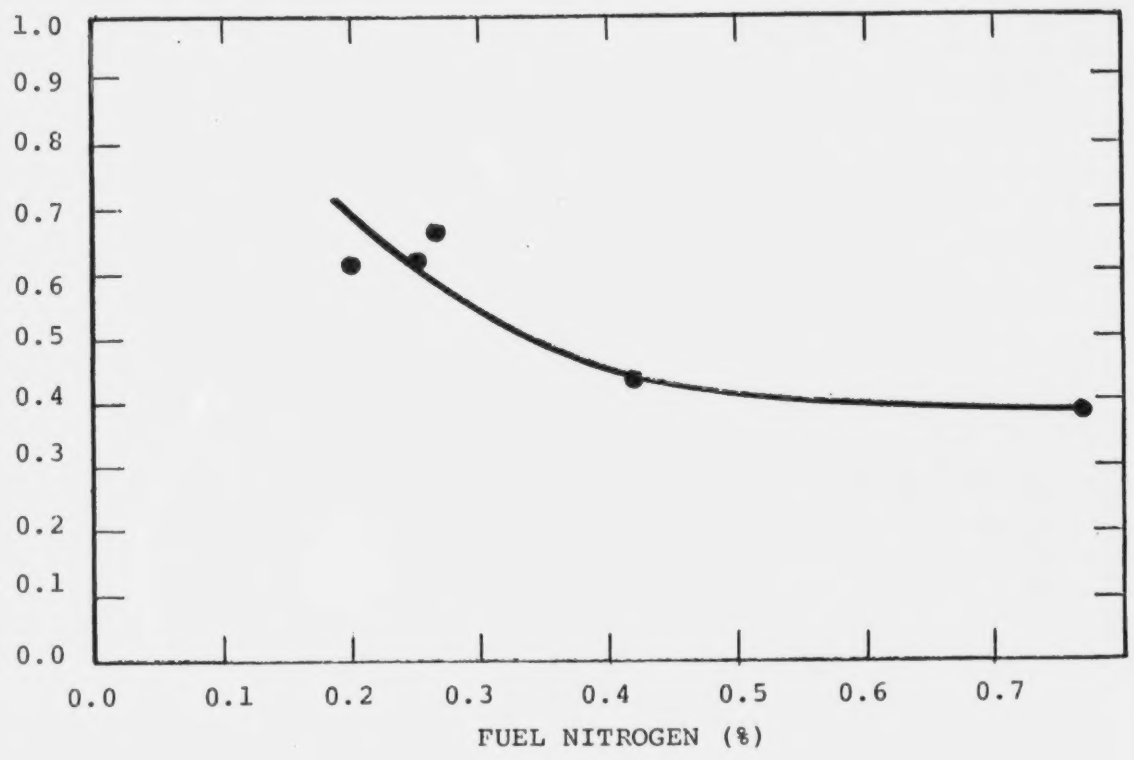

Figure 9-23. Fuel nitrogen conversion as a function of fuel nitrogen content.*

*From Item 5, p.62 in reference list

However, projected EPA standards will be considerably lower, as shown by the following table (4,p.36):

	NO_x, ppm (v)	
	1980	1985
Coal	200	100
Oil	150	90
Gas	100	50

Control of NO_x by Boiler Modifications--
The use of boiler modifications to lower NO_x emissions will probably receive the initial emphasis since these methods are the least costly. Design changes which lower the flame temperature and reduce oxygen availability result in lower NO_x formation. These modifications can be one or more of the following:

- Two-stage combustion
- Low excess air firing
- Flue gas recirculation

Several sources have somewhat differing opinions on the effects of these techniques, as shown in TABLE 9-3.

Therefore, it is concluded that boiler modification can result in a reduction in NO_x emissions of between 40 and 50 percent.

The following table (1, p.21) (3, p.95) gives the expected emissions of NO_x for both uncontrolled and controlled boiler conditions.

TABLE 9-3. NO_x REDUCTION WITH BOILER MODIFICATIONS

		NO_x Reduction, %		
	Ref.*	Coal	Oil	Gas
Staged Combustion	3	34	16	25
	4	40	40	55
	6	35	40	50
Approximate Average		35	35	45
Low Excess Air	3	22	20	24
	4	20	20	20
	6	25	33	33
Approximate Average		20	25	25
Combined Staged Combustion and Low Excess Air	3	38	30	42
	4	40	35	50
	6	60	73	90
Approximate Average		40	40	50
Flue Gas Recirculation	3	15	17	40
	4	NR	20	60
	6	33	33	33
Approximate Average		25	25	45

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*Numbers refer to items in the reference list
Item 3, p.95: ranges in the article have been averaged for this table.
Item 4, p.35: source for the article was EPA report 650/2-74-066.
Item 6, p.III-55.

	Expected NO_x Emissions		
	Coal	Oil	Gas
Uncontrolled Combustion, ppm	500	280	200
With Boiler Controls, ppm	370	150-210	85-110
NO_x Reduction with Boiler Controls, %	26	25-46	45-57

Therefore, it may be concluded that boiler modifications are sufficient to meet the most stringent present standards for NO_x emissions (perhaps with coal being a borderline case). These modifications also appear to meet the projected 1980 goals for oil and gas, but not for coal. However, none of the projected 1985 goals can be met solely by boiler modifications.

Control of NO_x by Reducing Liquid Fuel Nitrogen Content--
As previously mentioned, reducing the fuel nitrogen content will result in lower NO_x emissions. This can be accomplished by hydrodenitrogenation of the coal-derived liquids such as tars and tar oils. These liquids are expected to contain about the same percentage of nitrogen as the feed coal (7,p.124). The hydrodenitrogenation scheme is shown in Figure 9-24 (8,p.5).

Hydrodenitrogenation is accomplished by hydrogenolysis of organonitrogen compounds to ammonia and the corresponding hydrocarbon as shown below (8,p.3):

$$C_wH_xN_y+1/2(z+3y-x)H_2 \rightarrow yNH_3+C_wH_z$$

Petroleum oils are usually hydrotreated at a temperature in the range of 600-800°F, a pressure in the range of 300-4,000 psig and a hydrogen to oil ratio in the range of 300 to 15,000 SCF per barrel. Studies by ARCO (Atlantic Richfield Company) and Wan on COED oil showed that the degree of nitrogen removal is a function of hydrogen partial pressure, temperature, space velocity, and

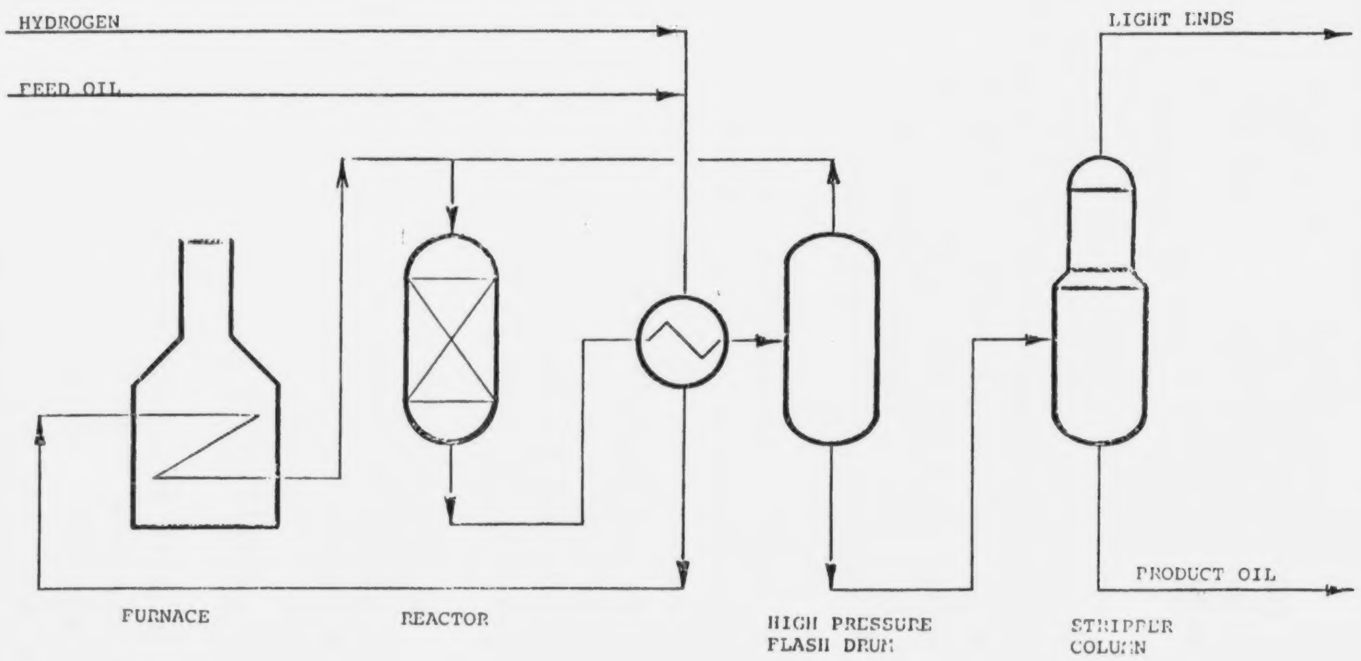
HYDROGEN
FEED OIL
LIGHT ENDS
PRODUCT OIL
FURNACE
REACTOR
HIGH PRESSURE FLASH DRUM
STRIPPER COLUMN

Figure 9-24. Typical hydrotreater unit.*

*From Item 8, p.5 in reference list

catalyst particle size. The removal is favored by higher pressures and temperatures and by lower space velocities and catalyst particle sizes. The effects of space velocity and pressure on denitrogenation are shown in Figures 9-25 and 9-26. A slight improvement in nitrogen removal can be effected by increasing the hydrogen recirculation ratio. Catalyst pore diameter appears to have no effect on denitrogenation. These studies have shown that up to about 80 percent nitrogen removal can be attained at pressures of 2,000 to 3,000 psig and temperatures of 750 to 800°F (8,p.14-21).

Similar studies by Satchell have shown that about 75 percent nitrogen removal can be attained at a pressure of 1,000 psig and a temperature of 750°F. Presumably over 80 percent removal could be achieved at higher pressure and/or for temperature (8, p.140). Figure 9-27 presents typical data from the investigations.

An example will serve to illustrate the benefits which accrue when liquid fuel denitrogenation is employed. An oil fired boiler with 2.7 percent oxygen in the flue gas (corresponding to about 15 percent excess air) which burns fuel containing 0.4 percent nitrogen will have about 280 ppm NO_x in the flue gas. Assuming 45 percent conversion of the fuel nitrogen to NO_x, the relative contributions of the different NO_x mechanisms are:

	ppm (v)
NO_x (Fuel at 45 percent Conversion)	207
NO_x (Thermal)	73
Total	280

If 80 percent of the fuel nitrogen is removed by the techniques discussed, the residual nitrogen content will be 0.08 percent;

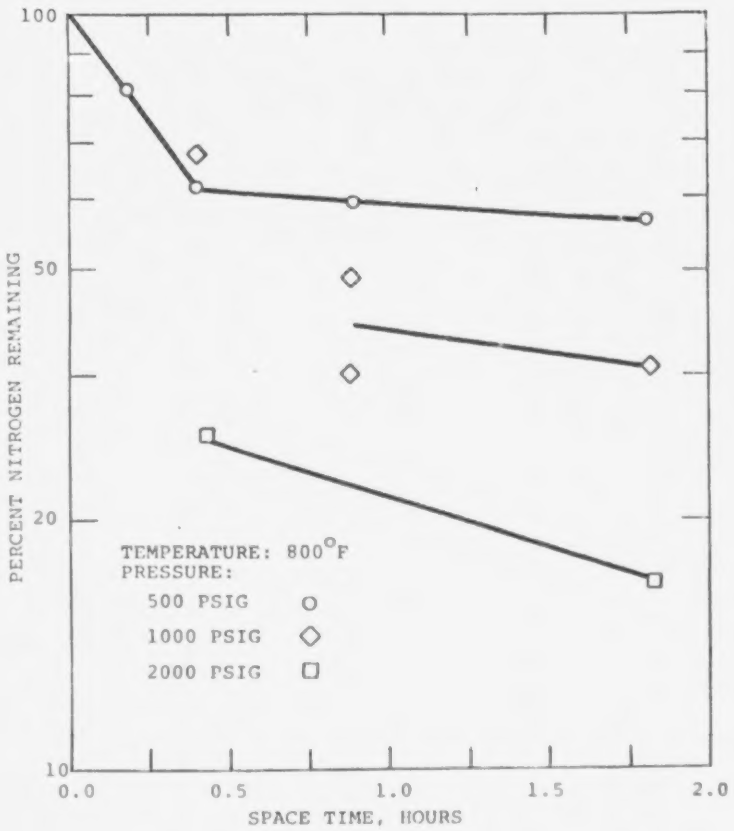

Figure 9-25. Effect of space time on denitrogenation of Anthracene Oil.*

*From Item 8 in reference list.

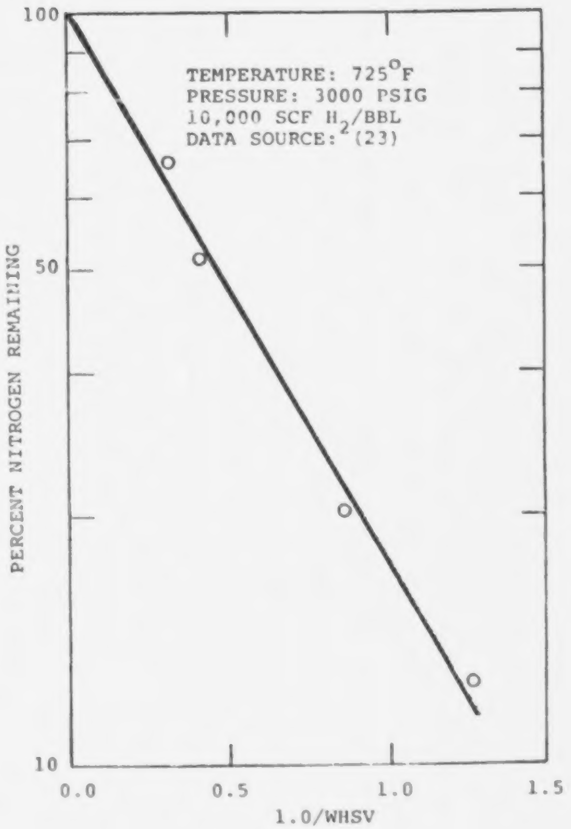

Figure 9-26. Effect of weight hourly space time on denitrogenation of COED oil.*

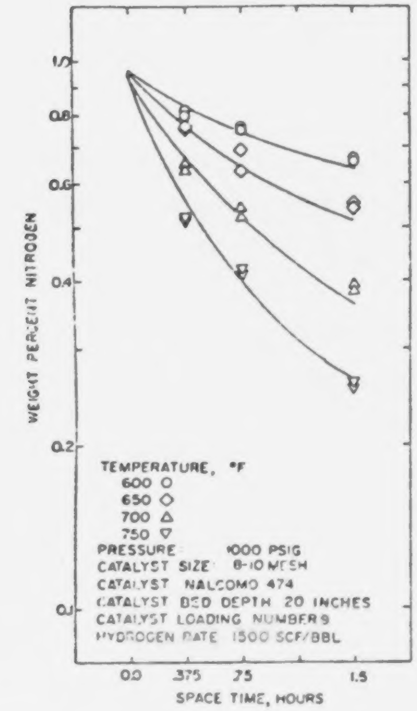

Figure 9-27. Effect of space time and temperature on denitrogenation.*

*From Item 8 in reference list

however, about 70 percent of the residual will be converted to NO_x. The resulting NO_x emissions become:

	ppm (v)
NO_x (Fuel at 70 percent Conversion)	65
NO_x (Thermal)	73
Total	138

If the original fuel contains twice as much nitrogen (0.8 percent) it is clear that a 90 percent denitrogenation efficiency is needed to achieve the same results.

These calculations are in good agreement with the data presented in Figure 9-28. In this example the NO_x emissions from an oil fired boiler were reduced about 50 percent, sufficient to meet the present most stringent standard and the projected 1980 goal. However, fuel denitrogenation alone will not be sufficient to meet the projected 1985 goal of 90 ppm NO_x for oil fired units. For this goal a combination of boiler modifications, to reduce NO_x emissions by 30 to 40 percent, and fuel denitrogenation might suffice:

	NO_x, ppm (v)
0.7 x 138 =	97
0.6 x 138 =	83

Control of NO_x (and SO_2) by Flue Gas Denitrification Processes--Currently, there are some 48 flue gas denitrification processes, both dry type and wet type, reported in the literature. A significant amount of information is reported on 42 of them.

For a variety of reasons, including simplicity, more favorable economics, and the fact that most existing power plants in Japan operate with low sulfur oil as fuel with small amounts of

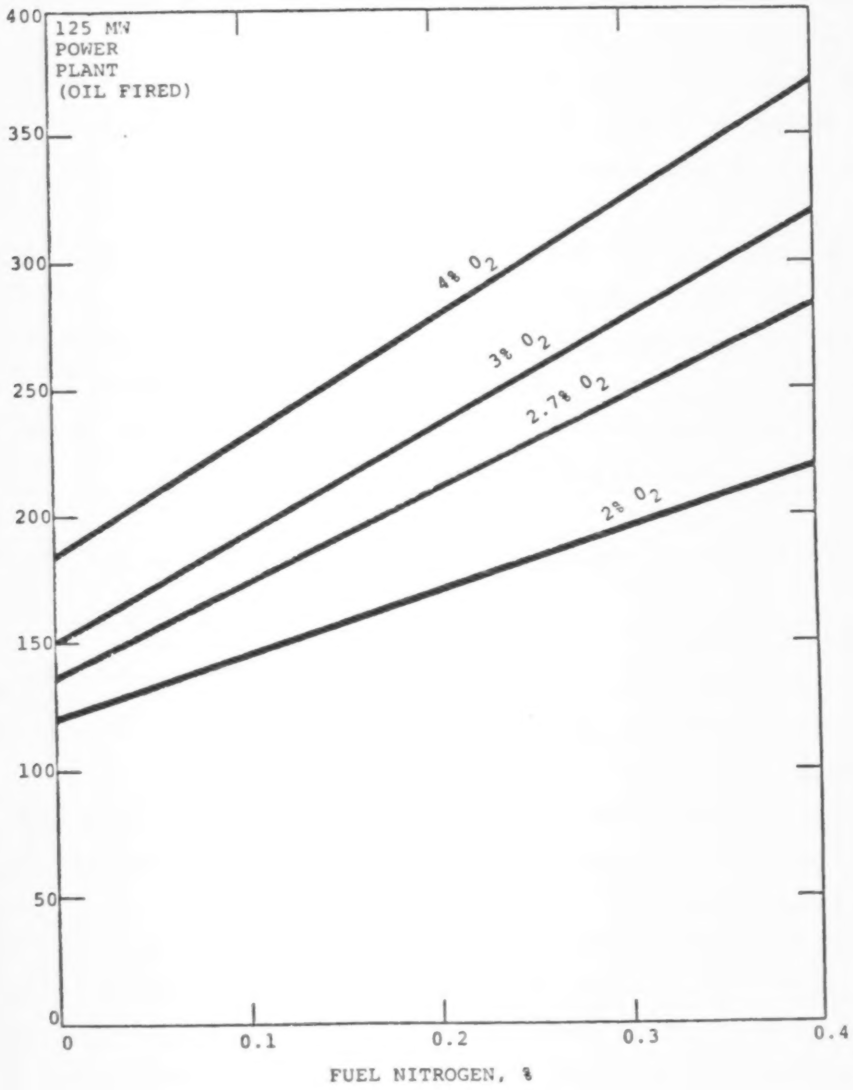

Figure 9-28. Effect of fuel nitrogen on NO_x emissions.*

*From Item 2, p. 13 in reference list

particulate and SO_2 flue gas emissions, the dry NO_x-only removal processes seemed the most promising and were developed first. The initial research for most of the wet processes was begun later and, hence, the dry processes as a group have been more extensively tested and are more commercially acceptable.

Although there are many different types of dry and wet processes, in most cases the dry processes have the following advantages over the wet processes:

- Lower projected total capital investment and lower annual revenue requirements
- Simpler process with few equipment requirements
- Higher NO_x removal efficiency (over 90 percent)
- More extensive tests in large oil and/or gas fired boilers
- No waste stream generation

However, dry systems also have the following disadvantages:

- More sensitive to inlet particulate levels
- Requirement for ammonia from either an energy-sensitive source (natural gas) or more expensive coal gasification methods
- Possible emission of ammonia, ammonium sulfate, and ammonium bisulfate. Precipitation of the latter two may foul downstream equipment
- Relatively higher reaction temperatures are required (350-400°C), which means that the system must be located in the power generation cycle before the air preheater or the temperature must be attained by auxiliary heating after the preheater

The most critical of these disadvantages, particularly for the U.S. utility industry with its heavy reliance on coal for power generation, is the sensitivity of these processes to inlet particulate levels. However, major research is now underway to develop methods so that dry systems can handle flue gas with high particulate loading.

Another disadvantage of the dry, selective catalytic reduction (SCR) processes is that the ideal catalyst location may be in the region between the economizer outlet and the air preheater inlet and, hence, the process is intimately involved in the power generation cycle. Therefore, if problems of operating these processes occur, the adverse impacts on the basic utility operations may be greater.

In addition to the above-mentioned disadvantages, the long-term supply of ammonia for these dry NO_x removal processes is a potential problem. Ammonia is the reducing agent for converting NO_x to molecular nitrogen for the SCR processes, which comprise nearly all of the dry processes and about half of all the NO_x removal processes, and the selective non-catalytic reduction process. With an NH_3:NO_x mol ratio of about 1:1, a single 500 MW coal-fired power plant (600 ppm NO_x in the flue gas) would consume about 5950 tons per year of liquid anhydrous ammonia. In view of the continuing increase in the world's demand for ammonia and ammonia-based fertilizers, the availability of ammonia for larger numbers of these dry NO_x removal units warrants concern and further investigation.

The wet NO_x removal processes have certain general advantages and disadvantages as compared with the dry systems. The major advantages include:

- Simultaneous SO_2 and NO_x removal may be a potential economic advantage
- Relative insensitivity to flue gas particulates
- Higher SO_2 removal (over 95 percent)

On the other hand, the major disadvantages of these wet systems include:

- More expensive processes due to the low solubility of NO_x in aqueous solutions and due to more extensive equipment requirements
- Formation of nitrates and other potential water pollutants
- Formation of low-demand byproducts
- Flue gas reheat required (however, if a wet SO_2 removal system were used in series with a wet removal system for NO_x only, the reheat would already have been incorporated into the design)
- Only moderate NO_x removal
- Application of some processes may be limited to flue gas with high SO_2:NO_x ratio

The two primary disadvantages of the wet systems are the high capital and operating costs and the formation of nitrate-containing wastewater. The formation of nitrate salts in most of these processes requires removal of these salts from the effluent by either evaporation or biological treatment.

Another inportant factor which must be considered in the application of certain wet NO_x removal processes is the minimum SO_2:NO_x ratios required. It may not be feasible to operate wet NO_x removal processes using flue gas from Western U.S. coals, which characteristically possess low amounts of sulfur, since sufficient SO_2:NO_x ratios may not be achieved for adequate NO_x removal (2,pp. xvi-xviii).

Of the 42 processes screened, the following 8 were selected for further study (2,p. xxviii):

	Type of Process (Classification)
UOP Shell Copper Oxide	Dry Simultaneous SO_2-NO_x (Selective catalytic reduction)
UOP Shell Copper Oxide	Dry NO_x only (Selective catalytic reduction)
Hitachi Zosen	Dry NO_x only (Selective catalytic reduction)
Kurabo Knorca	Dry NO_x only (Selective catalytic reduction)
Moretana Calcium	Wet simultaneous SO_2-NO_x (Oxidation-absorption-reduction)
Ishikawajima-Harima Heavy Industries	Wet simultaneous SO_2-NO_x (Oxidation-absorption-reduction)
Asahi Chemical	Wet simultaneous SO_2-NO_x (Absorption-reduction)
MON Alkali Permanganate	Wet NO_2 only (Absorption-oxidation)

In a 1977 study conducted by Pullman Kellogg for a client, the overall conclusion was reached that dry NO_x removal processes are definitely superior to wet processes. For this report the same viewpoint is adopted. Two of these dry processes will be described as illustrative examples of the type.

UOP Shell Process--The Shell copper oxide process that is licensed by Universal Oil Products (UOP) may be used to remove only NO_x from flue gas or to remove both NO_x and SO_2.

For removal of NO_x only, as shown in the block flow diagram, Figure 9-29, the flue gas leaves the boiler economizer, ammonia is injected and the mixed gas stream enters the reactor. The fixed bed in the reactor is composed of copper supported on special aluminum oxide. The copper first oxidizes to CuO and then reacts with the SO_2 in the flue gas:

$$SO_{2(g)} + 1/2\ O_{2(g)} + CuO_{(s)} \quad CuSO_{4(s)}$$

The copper sulfate acts as a catalyst for the reaction of NO_x with ammonia and the consequent reduction of the NO_x to nitrogen and water:

$$6NO_{(g)} + 4NH_{3(g)} \longrightarrow 5N_{2(g)} + 6H_2O_{(g)}$$

$$6NO_{2(g)} + 8NH_{3(g)} \longrightarrow 7N_{2(g)} + 12H_2O_{(g)}$$

The treated flue gas leaves the reactor and flows through the air heater, particulate removal and flue gas desulfurization equipment and is finally exhausted to atmosphere. The copper in the reactor bed becomes fully converted to copper sulfate to act as the catalyst for NO_x removal. In this form the catalyst has no effect on the SO_2 in the flue gas and the SO_2 passes through the reactor unchanged.

The UOP/Shell Process can be designed for simultaneous NO_x-SO_2 removal by adding a regeneration step, wherein part of the copper sulfate is converted to CuO by steam-diluted hydrogen rich gas and returned to the reactor bed while the rich SO_2 stream is sent to sulfur recovery or other disposal.

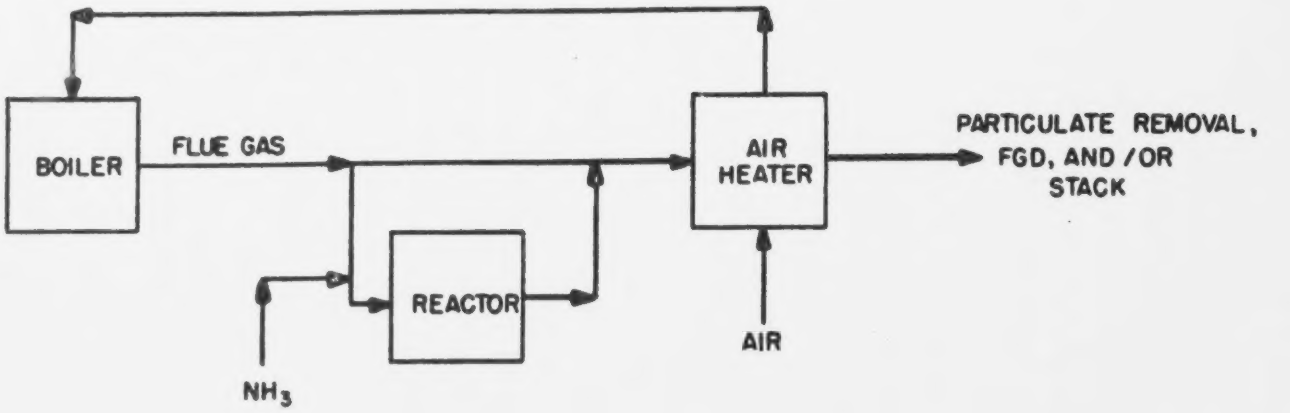
BOILER
FLUE GAS
NH$_3$
REACTOR
AIR
HEATER
AIR
PARTICULATE REMOVAL,
FGD, AND /OR
STACK

Figure 9-29. UOP/Shell NO$_x$ removal process.
(From Item 2 in reference list)

NO_x removal of at least 90 percent is expected with beds 4 meters long. Removal efficiencies of 95 to 97 percent with beds 6 meters long and 99 percent with beds 7 meters long have been achieved during prototype-scale testing.

All the reactions are exothermic. By locating the flue gas treatment section upstream of the air heater, the heats of reaction may be recovered together with the fan compression heat.

The Shell International Petroleum Company began development work in the early 1960's and in 1967 built a unit of 0.2 to 0.3 megawatt (MW) equivalent capacity for simultaneous NO_x-SO_2 removal at Pernis, The Netherlands to process flue gas from combustion of high sulfur fuel oil. Successful operation for 4 years and more than 20,000 regeneration cycles proved the feasibility of the process. In 1973 an NO_x-SO_2 unit of 40 MW equivalent capacity was started at Showa Yokkaichi Sekiyu (SYS), Japan on a flue gas from an industrial boiler containing 2,500 ppm SO_2 and has been operating successfully. A pilot plant of 0.6 MW equivalent capacity has been operating successfully for 3 years at the Tampa (Florida) Electric Company (TECO) on flue gases from combustion of coal containing 3.5 percent sulfur.

The UOP Process Division holds the worldwide (except for the Far East) licensing rights for the Shell process.

Operating parameters, capital investment and operating requirements reported (2) for a 500 MW installation are shown in TABLE 9-4.

The NO_x removal efficiency is reported to be independent of the

TABLE 9-4. UOP/SHELL NO_x CONTROL SYSTEM TREATING FLUE GAS FROM GENERATION OF 500 MEGAWATTS

Flue gas at 400°C, Ncu.m/hr	1,582,000
SO_2 inlet, ppm	2,580
NO_x inlet, ppm	634
NO_x outlet, ppm	63
NO_x removal	90%
Reactors (4)	4m bed length, 8.7m diameter
Catalyst, kg/reactor	51,480
Pressure drop across system, cm water	23
Capital investment (1977 Midwest U.S.)	$15,500,000

Operating Requirements

Ammonia, STPD	21
Power, KWH/day	77,800
Steam, STPD	13.2
Labor, Manhours/day	5
Maintenance, per year	$194,000
Heat credit, MBtu/hr	26.5
Operating Cost, mills/KWH	1.4

NO_x concentration in the inlet gas. The NO_x reaction is first order and the reaction rate is directly proportional to the NO_x concentration. Removal rate increases as the reactor bed length increases.

NO_x removal data from tests at pilot plants and the prototype scale unit (40 MW equivalent) at SYS are shown in Figure 9-30. The acceptor used was completely in the sulfated form. The results are plotted as a function of bed length, since conversion fluctuated with time on stream, design of reactor internals and operating conditions. NO_x removal efficiencies of 95 to 97 percent with beds 6 meters long and 99 percent for beds 7 meters long were obtained during tests.

The parallel-passage reactor has been shown capable of operating at normal conditions with full particulate loading of 10 grains or more per standard cubic foot. Initial testing with full flyash loading to the reactor at the Pernis unit showed no deterioration in performance. However, at SYS the high V- and Na-containing flyash gradually fouled the reactor internals, forcing runs to be limited to operating periods of 1 to 2 months. The effect of the fouling on acceptor performance is shown in Figure 9-31.

The higher particulate loading with coal fired flue gas at TECO of 10 grains or more per standard cubic foot caused a decline in performance after only days of operation. As a result of this problem, a procedure was developed and implemented at this pilot plant which provided in situ cleaning of reactor internals during normal operation. This technique allowed stable performance with high particulate loadings. Pressure drop across the reactor and desulfurization were not affected.

Testing at TECO also demonstrated that chlorine and chlorides in

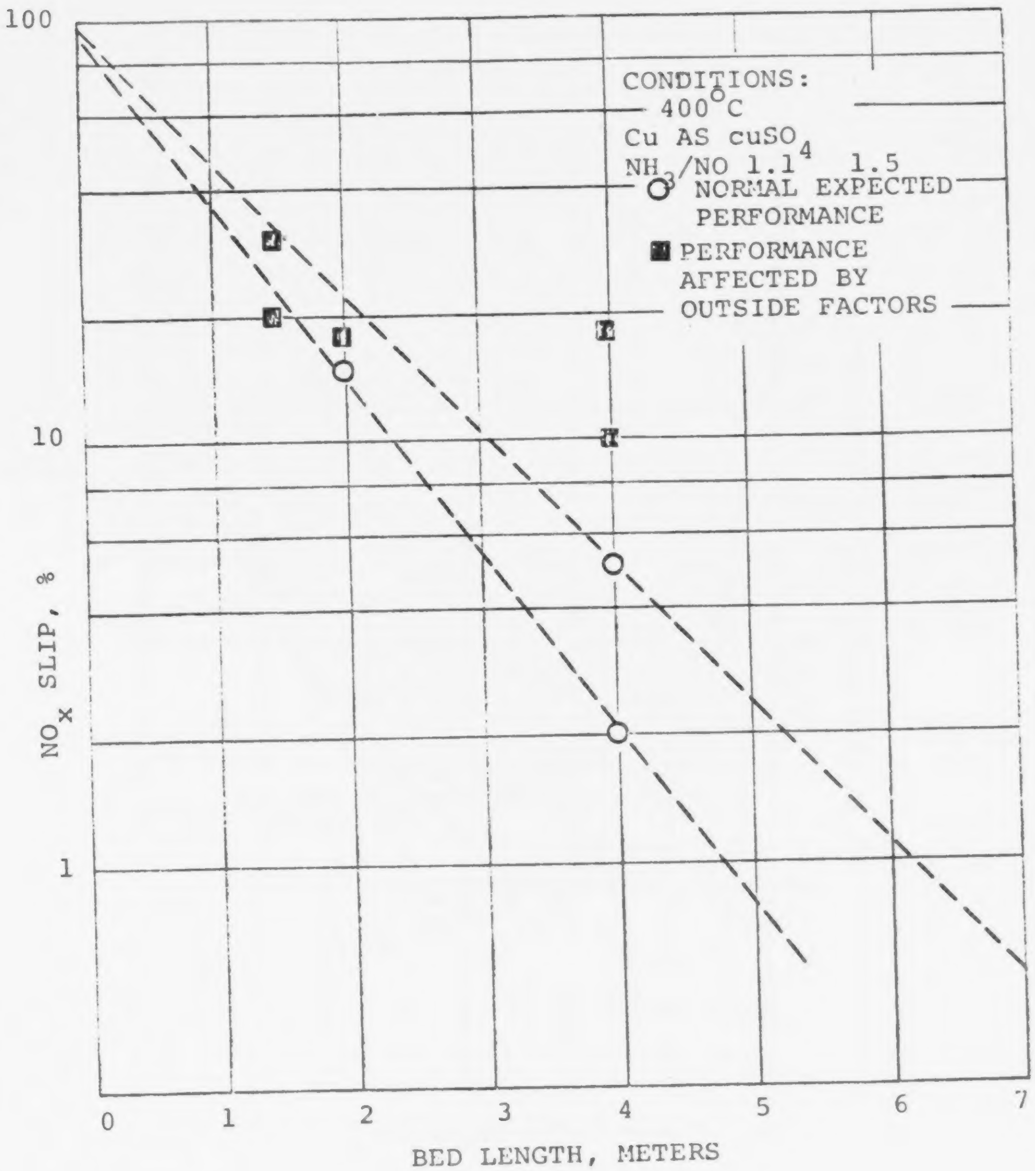

Figure 9-30. Unconverted NO$_x$ as a function of catalyst bed length for UOP process.*

*From Item 2 in reference list

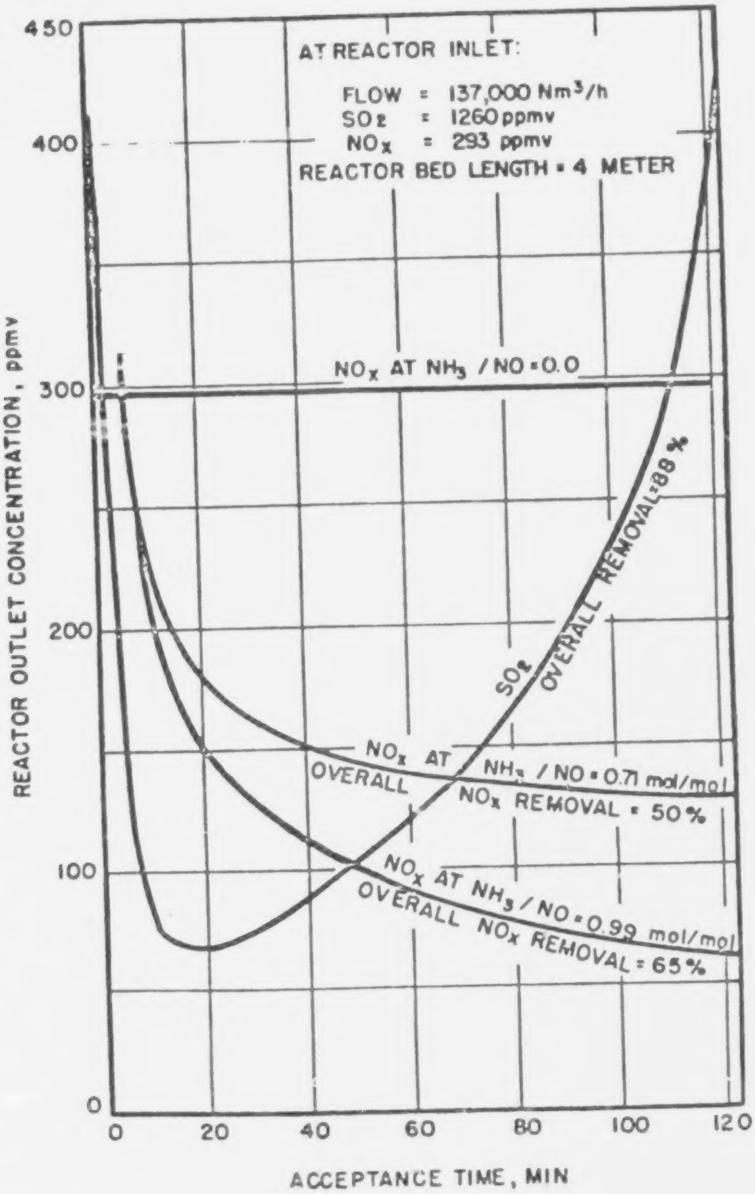

Figure 9-31. Performance of Shell reactor at SYS: instantaneous SO_2 and NO_x slip.*

*From Item 8 in reference list

flue gas had no adverse effect upon the acceptor performance. The loss of copper was negligible.

The use of an alumina base for the copper sulfate catalyst is questionable since competing processes appear to be changing to alternate supports that are supposedly more stable. It is recognized that UOP refers to their catalyst as being on a "special" alumina support.

The UOP/Shell process has these advantages:

- Achieves NO_x removal efficiencies of 90 percent or more
- Has been applied to flue gas from commercial oil-fired boilers
- Is a slight modification of a commercially available flue gas desulfurization system
- Operates with full particulate loading of 7 to 10 or more grains per standard cubic foot. (Note: After cyclones, loading should be 1.5 to 2 grains.)
- Claims less than 10 ppm by volume of ammonia in the treated flue gas. Pilot and prototype models have shown an average of 1 ppm in the flue gas
- Claims full turndown capability
- Requires no waste disposal for the system since no by-products are produced
- Although UOP considers information on materials of construction to be proprietary, statements are made that no unusual materials are used and that a service life of at least 15 years can be expected.

The disadvantage that must be cited is the lack of test data on flue gas from large scale coal fired units.

Hitachi Zosen Process--Hitachi Zosen (Hitachi Shipbuilding and

Engineering Compay, Ltd.) has developed an NO_x removal process in which dry, selective catalytic reaction of NO_x with ammonia occurs. Hitachi Zosen is now developing a catalyst and reactor design which permits treatment of flue gas with a high particulate loading. Therefore, the flue gas from a coal fired boiler may be fed directly to the reactor, upstream of the air heater, without any particulate removal treatment.

Ammonia is injected into the flue gas ahead of the reactor. In the reactor NO and NO_x are reduced to nitrogen and water by reaction with ammonia in the presence of a catalyst. Any excess ammonia is oxidized to nitrogen and water; therefore, Hitachi Zosen reports that there is no problem of excess ammonia in the flue gas leaving the system. The treated flue gas passes to the air heater for heating the air feeding the boiler. After the treated flue gas is cooled in the heat exchanger, the gas is sent to the particulate removal and desulfurization steps, and then leaves through the stack.

About 0.8 to 1.2 moles of ammonia are fed per mole of NO_x. The reaction temperature is 300 to 400°C. The area velocity (flow rate of gas per unit surface area of catalyst) is between 7 and 10 N cu.m/hr/sq.m. Catalyst surface area ranges from 550 to 600 sq.m/cu.m. NO_x removal efficiency is reported to be above 90 percent. The process is able to function with inlet particulate loadings up to 16 gr/N cu.m (about 7 grains per standard cubic foot). Hitachi Zosen claims that the pressure drop across the reactor is very low.

The catalyst is manufactured in the shape of units of honeycombs. These units are welded together for the particular size required. The flue gas passes parallel to the catalyst surface. The catalyst composition has not been revealed for proprietary reasons; however, Hitachi Zosen does state that it is constructed of common material. The expected catalyst life is reported to be one year.

Hitachi Zosen has been developing an NO_x removal process and catalyst since 1970. Five different catalyst series have been created and are known as NOXNON 100, 200, 300, 400, and 500 series. The 100 series is nonselective and for use with CO, hydrogen, and hydrocarbons as reductants. The other four series are for use with ammonia as the reducing agent. Series 200 is for treating "clean" flue gas, that is, gas which does not contain any significant amount of SO_x or particulates. The 300 and 400 series are resistant to SO_x. The 500 series is for use in gas with a considerable particulate loading. There have been over 21 different pilot plant tests since 1973 in developing these catalysts. The 500 series is still being tested and the results to date have been promising, even on gas containing 15 gr/Ncu.m(6.44 grains per standard cubic foot) of particulates and 300 ppm SO . The following pilot plant and bench scale units are now testing the NOXNON 500 series catalyst:

Source of flue gas (plant type)	Fuel	Test unit capacity, Nm^3/hr
Iron ore sintering	Heavy oil	5,000
Iron ore sintering	Heavy oil	200
Power	Coal	200
Glass melting furnace	Heavy oil	200

A commercial scale NO_x removal facility was constructed by Hitachi Zosen in the fall of 1975 for a petroleum plant with an oil fired boiler and flue gas treatment capacity of 440,000 Nm^3/hr. Another facility was completed in the fall of 1975 for a petroleum opeation with 350,000 Nm^3/hr flue gas capacity. Two more denitrification systems for steel manufacturing plants with

71,000 and 750,000 Nm^3/hr flue gas rates have been completed and are reported to be operating successfully.

The estimated capital cost for a denitrification unit to treat flue gas containing 300 ppm NO_x on a 250 MW plant is reported to be about $16/KW. The corresponding operating cost would be approximately 1.5 mills/KWH. These are assumed to be 1976 costs at a Japanese site.

The only major raw material required is ammonia. For a 250 MW plant with 300 ppm NO_x in the flue gas, an estimated 5 short tons/day of ammonia is required for NO_x removal. On the same basis, the energy consumption for denitrification is estimated to be:

Electricity	400 KW/day
Steam	4.1 Tons/day

No reheat fuel is necessary. The electrical requirement represents 0.2 percent of the total power output of the plant. The majority of the steam (2.9 tons/day) is used for ammonia vaporization, while the remaining 1.2 tons/day is for soot blowing.

With no particulate scrubbing before the NO_x removal system, soot blowing of the air heater and the reactor are required. For pilot plants, the reactor has required washing out of deposits once every 2 months. Hitachi Zosen states that this operation requires only a short time to complete. The technical support necessary for operation of this NO_x removal system should be similar to other dry systems.

There is no information available at present on the sensitivity of the NO_x removal efficiency to factors such as inlet gas composition and various operating conditions. Specifically, there is no comment on any possible adverse effects from chloride in the flue gas from a coal fired boiler. These effects might include interference of NO_x removal efficiency or deterioration of catalyst.

The composition of the catalyst is not known. The exact cost is not known, though the estimated cost for the initial investment in catalyst necessary for a 250 MW plant with flue gas containing 300 ppm NO_x is given by Hitachi Zosen as $1.7MM (1976 cost and assuming 300 yen = $1).

There are few available facts on wash frequency and other washing requirements on the reactor treating the flue gas directly from a coal fired boiler with no particulate removal. More information should be accessible as the pilot plant tests progress.

Maintenance, operating and technical support requirements have not been published. The major material of construction is reported to be carbon steel.

There are no pollutants removed, other than NO_x , by the Hitachi Zosen process. There is no byproduct from the process.

Excess NH_3 at the outlet of the system is a potential problem; however, Hitachi Zosen claims NH_3 leaving the system is very low and creates no pollution problem. Some type of disposal for the reactor deposit washing solution would be essential. Also, the catalyst is disposed of and not reclaimed.

The Hitachi Zosen process has these advantages:

- Achieves NO_x removal efficiencies of 90 percent or more
- Has been applied to flue gas from commercial oil fired boilers

- Operates with full particulate loadings of 7 or more grains per standard cubic foot
- Probably contains less than 10 ppm by volume of ammonia in the treated flue gas
- Claims full turndown capability
- Requires no waste disposal for the system since no by-products are produced
- Claims carbon steel to be the major material of construction

The disadvantages are principally technical and reflect data gaps regarding effects of inlet gas composition and variations in operating conditions.

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Processes and Techniques for Control of Sulfur Dioxide

When sulfur is a component of fuels fed to combustion equipment, its oxidation leads to the formation of sulfur dioxide (SO_2) and sulfur trioxide (SO_3). Normally about 98 percent of the sulfur leaving the equipment will be SO_2.

In coal, sulfur is present in three forms:

- pyritic sulfur, (FeS_2)
- organic sulfur, where the sulfur is chemically bound in the coal molecules
- sulfate sulfur, $FeSO_4$ or $CaSO_4$, normally less than 0.1 percent

In liquid fuels, sulfur is normally present in organic form. Tars and tar oils produced in low temperature gasification processes (e.g. Lurgi) normally contain appreciable amounts of sulfur although the total sulfur content will amount to only 50 to 80 percent of that of the feed coal (1, pp. 122-123).

Gaseous fuels such as natural gas contain very little sulfur. However, miscellaneous waste gas streams produced in coal conversion facilities may contain appreciable amounts of sulfur as H_2S, COS, CS_2, and sulfur vapor.

Standards for sulfur dioxide emissions have become very stringent in the last few years. Projected future standards indicate that allowable maximum emissions of SO_2 will be even lower. The most stringent current standards for SO_2 from fuel burning equipment are the following:

	Lbs SO_2/MM Btu	
	>250 MM Btu/hr	<250 MM Btu/hr
Coal Burning	0.20 (Wyoming)	1.20 (Oklahoma)
Oil Burning		
Residual Oil	0.34 (N.Mexico)	1.00 (Illinois)
Distillate Oil	0.30 (Illinois)	0.30 (Illinois)
Gas Burning	0.13 (Montana)	0.13 (Montana)

Perhaps a few examples will help to put these standards into perspective. The lowest sulfur coals in the United States are generally found in the western states. The sulfur content of these coals is about 0.5 percent (minimum) and the higher heating value is about 10,020 Btu/lb. Burning this fuel without control will result in an SO_2 emission of about 1.0 lb/MM Btu which is higher by a factor of 5 than the most stringent standard for fuel burning equipment with a heat input greater than 250 MM Btu per hour. Therefore, it may be concluded that some type of sulfur control will be needed whenever coal is burned in these fairly high capacity ranges.

A typical tar from a gasification facility has a higher heating value of about 16,500 Btu per pound (1, p.121). Its sulfur content will be on the order of 60 percent of that of the feed coal. In order to meet a standard of 0.34 lb SO_2/MM Btu, the tar or tar oil can contain only 0.28 percent sulfur, which corresponds to a coal containing about 0.47 percent sulfur. Therefore, for this standard, tars and oils from only the lowest sulfur coals could be burned without controls. However, when meeting a standard of 1.0 lb SO_2/MM Btu, the tar could contain 0.82 percent sulfur, which corresponds to a coal containing about 1.37 percent sulfur. Thus quite a few coals would produce tars which could qualify for combustion without controls.

It is difficult to generalize regarding combustion of gas streams because of the wide range of heating values and sulfur contents of these fuels. A typical low Btu gas may have a higher heating

heating value of about 200 Btu per standard cubic foot (SCF) (1, p.24) (2, p.315) or about 3,030 Btu/lb (molecular weight = 25). To meet a standard of 0.13 lb SO_2/MM Btu, such a gas can have a maximum sulfur content of about 0.02 percent by weight or about 200 ppm by weight. On the other hand, a high Btu gas (1,020 Btu/SCF, HHV) can have a maximum sulfur content of about 0.15 percent by weight or 1500 ppm by weight and still qualify for combustion in boilers. However it is very unlikely that this valuable product would be used in such a manner. More likely, a high Btu gas (SNG) would be delivered to the pipeline and in this case the maximum sulfur content permitted is 1/4 grain/SCF or about 4 ppm (2,p.3).

There is a broad spectrum of waste gas streams from coal conversion facilities which are potential candidates for combustion. Each stream, or combination of streams, will require analysis to determine the degree of sulfur control, if any, that is required for combustion.

Desulfurization of Liquid Fuels--

Liquid fuels produced as byproducts of coal conversion facilities will have appreciable sulfur contents. In order to reduce the sulfur content of these fuels to acceptable levels, hydrodesulfurization as practiced by petroleum refineries may be an attractive choice. Some of these processes are listed in TABLE 9-5.

The Gulf HDS process may be regarded as typical of the hydrodesulfurization schemes. A simplified flow diagram in is shown in Figure 9-32.

In this process, the heavy oil passes through a solids removal section which separates filterable solids, such as iron compounds. The oil feed is mixed with makeup and recycle hydrogen,

TABLE 9-5. IMPORTANT HYDRODESULFURIZATION PROCESSES*

LICENSOR	PROCESS	SCOPE OF APPLICATION
Chevron Research Co.	VGO Isomax	Vacuum gas oil and lighter feedstocks
	RDS Isomax	Whole crude, vacuum gas oil, and vacuum tower bottoms
	VRDS Isomax	Vacuum tower residuum
Cities Service R&D Co. and Hydrocarbon Research Inc.	H-Oil	Residual oil
ESSO Research & Engineering Co. and Union Oil of California	Go-Fining Residfining	High boiling virgin and cracked gas oil Atmospheric tower residuum
Gulf Research & Development Co.	Gulf HDS Type I and Type II	Atmospheric tower residuum
	Gulf HDS Type III	Atmospheric tower residuum
	Heavy Gas Oil Gulfining	Virgin or cracked heavy gas oils
Institute Francais de Petrole	IFP Vacuum Gas Oil HDS	Vacuum gas oil
	IFP Resid HDS	Atmospheric tower residuum
Standard Oil Co. (Indiana)	Resid Ultrafining	Vacuum residuum
	VGO Ultrafining	Vacuum gas oil
Universal Oil Products Co.	RCD Isomax	Atmospheric reduced crude, deasphalted vacuum tower residuum
	Hydrobon Process	Light and heavy distillates

*From Item 2, p. XI-10, in reference list

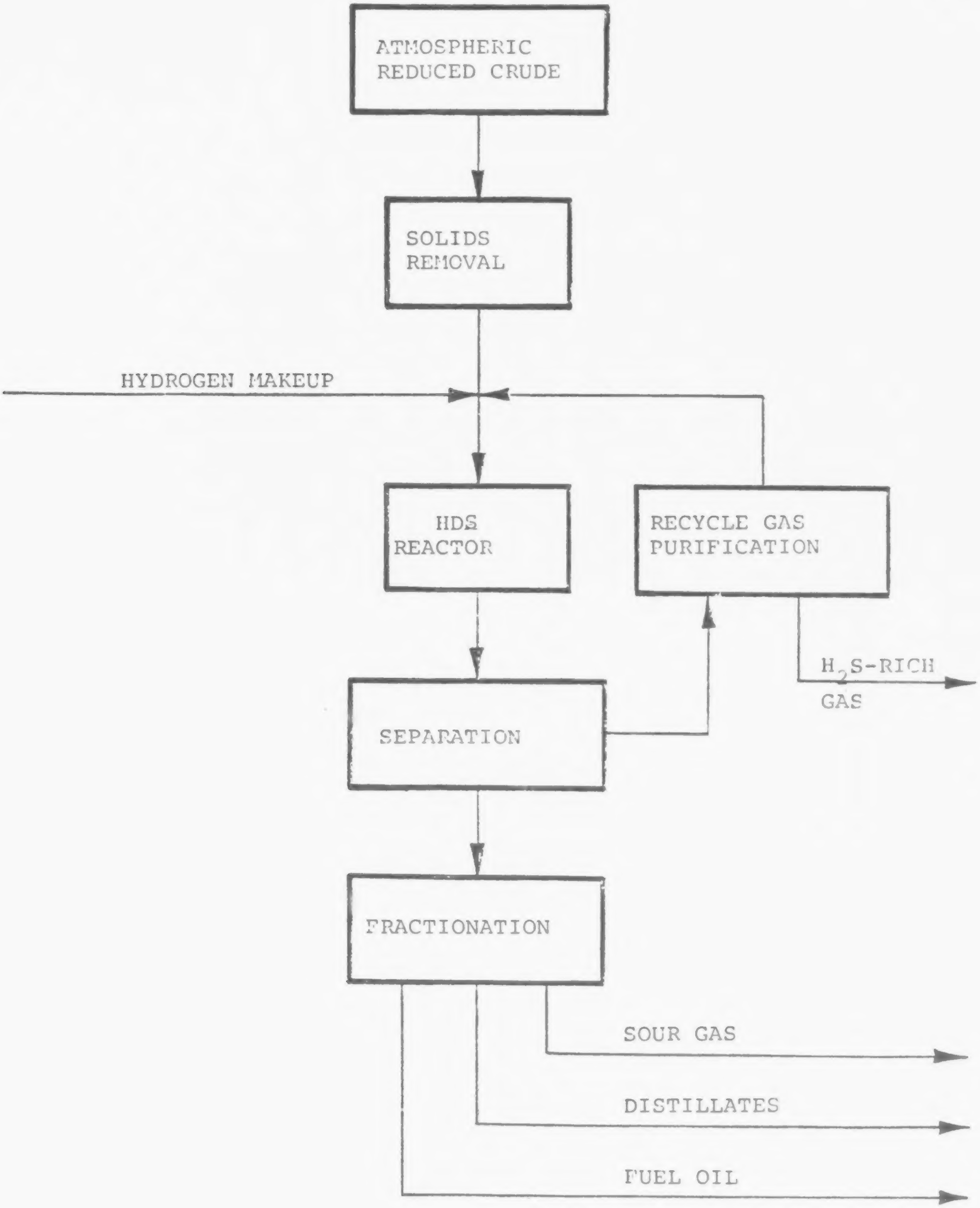
ATMOSPHERIC REDUCED CRUDE
SOLIDS REMOVAL
HYDROGEN MAKEUP
HDS REACTOR
RECYCLE GAS PURIFICATION
H_2S-RICH GAS
SEPARATION
FRACTIONATION
SOUR GAS
DISTILLATES
FUEL OIL

Figure 9-32. Typical flow scheme for a fixed bed hydrodesulfurization process.*

*From Item 2 in reference list

heated, and reacted over a fixed bed of catalyst at elevated pressure and temperature with evolution of H_2S. Specific operating conditions vary depending on the type of feedstock, the desired product, and the particular process. For residuum feed, the pressure may range upward from 71.2 kg/cm^2 (1,000 psig). Typical reaction temperatures are 399 to 454 degrees C. The reactor effluent is cooled and recycle gas is separated. Sulfur is removed from the recycle gas before it joins the reactor feed. Separator liquids flow to fractionation or stripping.

When distillates are fed to desulfurization processes, the solids removal step is unnecessary since these distillates are already purified liquids. The remaining processing steps are essentially the same as illustrated. However, with lower boiling distillates the catalyst quantity needed in the reactor is less and the pressure and temperature conditions are less severe for a given level of sulfur removal. For distillates, hydrotreating pressures may range upward from 29 kg/cm^2 (400 psig) and the temperatures may be between 371 and 427 degrees C.

An alternative system to the fixed bed reactor is used in the H-Oil process which was developed by Cities Service R&D and Hydrocarbon Research, Inc. This process uses an ebullating bed of catalyst through which the oil and hydrogen flow. The ebullating bed (similar to a fluid bed but with liquid as the dispersing medium) process is competitive with fixed bed systems primarily when residuums are fed. Advantages of the ebullating bed reactor over a fixed bed reactor in such a system are:

- Solids contained in the feed pass through the ebullating bed and do not cause the plugging that occurs with fixed beds.

- Catalyst can be added and removed from the ebullating bed reactor while it is in operation, thus avoiding the necessity for shutting down when catalyst activity is too low
- In general, smaller catalyst particles can be used in ebullating beds. These show higher reactivity for residuum desulfurization.

A sulfur removal efficiency of about 90 percent can be achieved with these methods (2,pp. XI-3-13).

Desulfurization of Coal Prior to Combustion. The Meyers Process--

The Meyers process, being developed by TRW, Inc., is reported to be effective in removing pyritic sulfur from coal. In this process, the pyrites in the coal react with ferric sulfate in a solution containing ferric and ferrous sulfates and sulfuric acid. The ferric ion is continuously regenerated by reaction of oxygen and ferrous ion. The elemental sulfur product is extracted with an organic solvent. The iron product from the pyrites is removed as solid ferric and ferrous sulfates. A block flow diagram of the basic Meyers process is shown in Figure 9-33.

Coal ground to less than 100 mesh is mixed with recycled leach solution and is then pumped from the mixing vessel to one of 10 reactor vessels where the slurry is contacted with oxygen at about 150°C. The pyritic sulfur is 90 percent converted to elemental sulfur and sulfate according to the following reactions:

$$FeS_2 + Fe_2(SO_4)_3 \longrightarrow 3FeSO_4 + 2S$$

$$FeS_2 + 7Fe_2(SO_4)_3 + 8H_2O \longrightarrow 15FeSO_4 + 8H_2SO_4$$

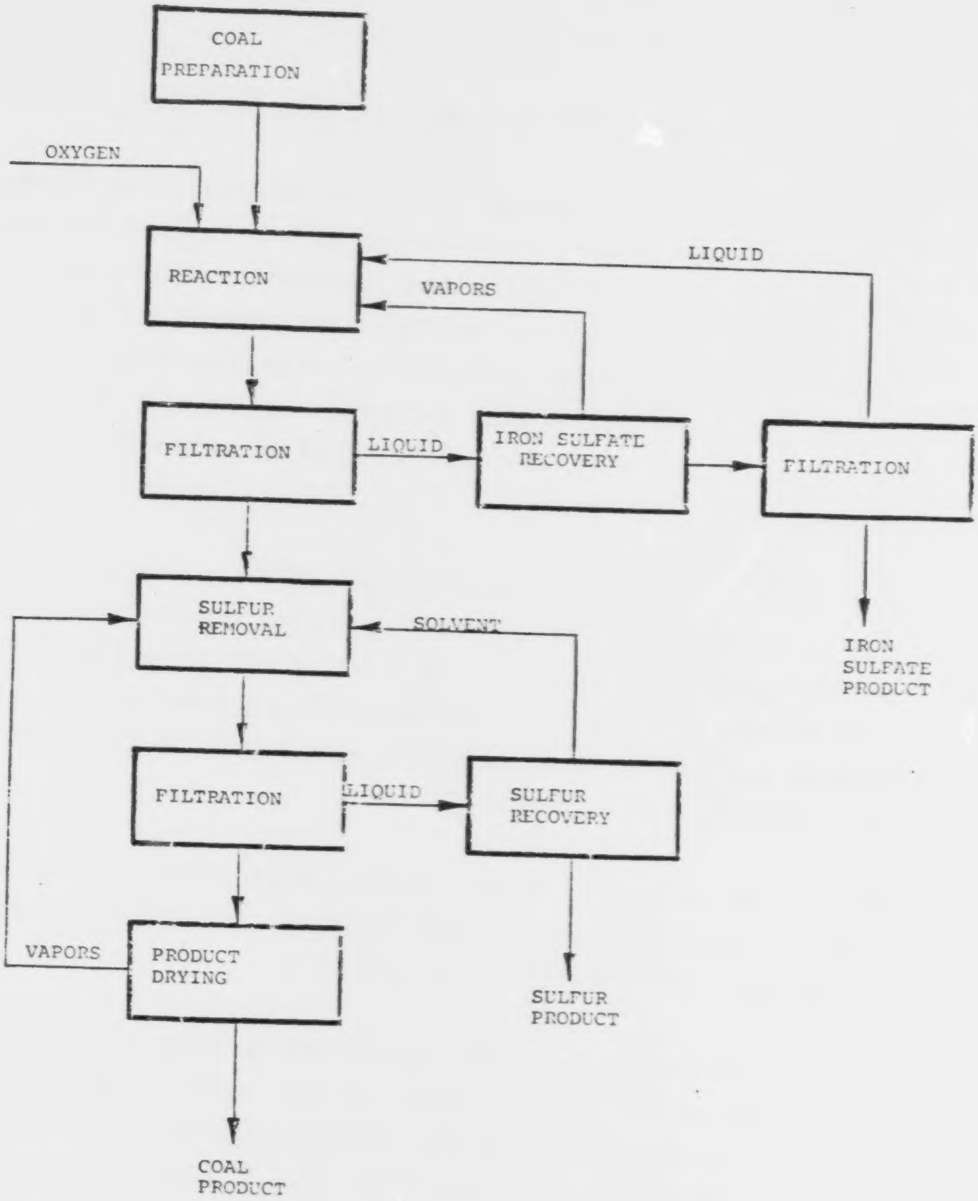
COAL PREPARATION
OXYGEN
REACTION
LIQUID
VAPORS
FILTRATION
LIQUID
IRON SULFATE RECOVERY
FILTRATION
IRON SULFATE PRODUCT
SULFUR REMOVAL
SOLVENT
FILTRATION
LIQUID
SULFUR RECOVERY
SULFUR PRODUCT
VAPORS
PRODUCT DRYING
COAL PRODUCT

Figure 9-33. Flow diagram of the modified Meyers process*
*From Item 3, pp. 4-23 in reference list.

$$4FeSO_4 + 2H_2SO_4 + O_2 \longrightarrow 2\,Fe_2(SO_4)_3 + 2\,H_2O$$

The net overall reaction for the leaching and regeneration steps may be considered as:

$$FeS_2 + 2.4\,O_2 \longrightarrow 0.2\,Fe_2(SO_4)_3 + 0.6\,FeSO_4 + 0.8\,S$$

The reaction slurry passes first to hydrocyclones, where about 60 percent of the liquid is removed and recycled to the reactors, and then to filters, where the remainder of the leach solution is removed and sent to iron sulfate recovery. The filter cake is reslurried in recycled naphtha to dissolve most of the elemental sulfur and then filtered. Water is separated from the filtrate by decantation and the sulfur-laden solvent flows to sulfur recovery.

The filter cake, containing about 25 percent moisture and 5 percent solvent, is partially dried under vacuum, where the sensible heat of the coal is sufficient to drive off the solvent and reduce the moisture to about 20 percent in the product coal. Vapors are condensed, water is separated by decantation and both are reused in the process.

In sulfur recovery, distillation separates the solvent from the product sulfur. Water and solvent in the overhead vapors are condensed, separated and recycled.

In iron sulfate recovery the filtrate from the reaction slurry is heated to about 130°C. Some of the water flashes and is used for heating in the reaction and sulfur recovery sections. The slurry of iron sulfates is filtered, the filtrate is recycled to the reaction section and the cake is sent to disposal.

TABLE 9-6 is an example of operation of the process (3), where a coal feed containing 3.92 percent total sulfur yielded a coal product containing 0.95 percent total sulfur, amounting to over 75 percent reduction in total sulfur. The feed and product contained 3.21 percent and 0.17 percent of pyritic sulfur, respectively, demonstrating pyritic sulfur removal of over 95 percent. Because of the reduction of the ash content of the coal product, the higher heating value increased by about 5 percent, to 12,747 Btu/lb. The thermal efficiency for this example of the process is 92.1 percent.

The Meyers process, as reported, is effective in removing pyritic sulfur from coal and has an attractive potential for treating, prior to combustion, those coals whose sulfur content is predominantly in the pyritic form.

It is of interest to note that the total of the forms of sulfur other than pyritic in the feed and in the product coal are virtually identical, indicating that the Meyers process does not significantly affect the other forms.

Desulfurization of Coal Prior to Combustion. Physical Cleaning--

Physical removal of pyritic sulfur is the most highly developed method technologically and potentially the lowest in cost.

The coals of the United States have highly variable characteristics by seam and by geographic location. Since coals vary so widely, coal cleaning processes are typically engineered for each coal source and designed with respect to the use to be made of the coal.

Coals are prepared by size reduction and subsequent particle sorting based upon particle size and density. The level of coal

TABLE 9-6. DESULFURIZATION OF COAL VIA THE MEYRS PROCESS

Basis: Processing 200,000 pounds per hour of dry coal

Ultimate Analysis of Feed, by Weight

C	68.53%	S-Pyritic	3.21%
H	3.85	-Sulfate	0.04
N	1.20	-Organic	0.67
Cl	0.08	Ash	20.86
		O	1.56

Sulfur Balance, Pounds per Hour

In Feed -	Pyritic	6,416
	Sulfate	80
	Organic	1,340
Total Input		7,836
Coal Loss		2
$FeSO_4$		1,828
$Fe_2(SO_4)_3$		1,780
Product Sulfur		2,438
In Product Coal to Boiler		125
In Net Product Coal -	Pyritic	298
	Sulfate	53
	Organic	1,242
	Elemental	70
Total Output		7,836

Thermal Efficiency

HHV of dry coal feed at 12,140 Btu/lb =	2,428 billion Btu/hr
HHV of net dry coal product at 12,747 Btu/lb =	2,235 billion Btu/hr
Thermal efficiency = Heat out/Heat in =	92.1%

quality improvements attainable is variable, being constrained by processing objectives, cost, processing technology, and coal characteristics.

Plants are designed to produce a product or products of definitive characteristics for one specific customer. The preparation plant is designed to remove the non-combustibles from the coal at the minimum practical operating cost and at the optimum practical yield. However, the ROM coal is prepared only to the extent that is necessary to make the product salable.

The technical limitations of the preparation process relate primarily to the very small component particles existing in coal. Many of these particles are residual structures of vegetation and minerals, generally irregular in shape. The pyrite particles in many coals are less than 1 micron (0.00004 inch) in their longest dimension. Particles smaller than 50 microns cannot be practically separated from each other, and separating them is usually inefficient. Larger particles, or those less homogeneous in composition, respond more readily to separation.

To be separable, impurity-containing particles must have masses greater than the pure coal particles. The difficulty in separating particles of less than 50 microns results from their slower response to the acceleration of gravity than larger particles: they literally float within the coal. Moreover, since most of the separation is done in water systems, there is the further complication that removal of the water from the small particles is significantly more difficult and more costly than removing water from the larger particles due to the smaller porosity of the smaller particles or of the combination of particles. Because of the technical difficulty in separating small particles, the separation costs increase as the particle size decreases. The processes which will remove more pyrite from the coal necessarily utilize smaller particle sizes and are considerably more costly. Accordingly, coal cleaned primarily for ash removal is cleaned with as large a particle size as is practical. It is for this reason that coal processing plants which were not designed for sulfur removal often do not function well as pyrite removers.

The economic limitations of coal preparation are varied and numerous. Cleaning of coarse coal is relatively simple and less costly than cleaning of the finer sizes. The fine coal portion in the raw coal feed has materially increased as mechanization of mining process has increased, thus adding considerably to cleaning plant costs. Wet cleaning units for fine coal are not expensive; it is the equipment necessary to dewater and dry the product that adds significantly to the cost. Clarifying the process water and thermal drying substantially increase plant capital investment. Yet many modern cleaning plants must contain this equipment in order to obtain the desired ash, sulfur, and moisture in the product and still recover the greatest amount of salable coal.

In 1974, 258 million one of a total of 603 million tons of U.S. coal produced were cleaned by wet methods (9,p.31). The following table (11,p.1-3) shows total U.S. coal production and the amount cleaned for the years 1964-1972.

PRODUCTION AND CLEANING OF U.S. COALS

	Millions of Tons								
	1964	1965	1966	1967	1968	1969	1970	1971	1972
Total Production	487	512	534	553	545	561	603	552	595
Quantity Cleaned	310	332	341	349	341	335	323	271	293
Percentage Cleaned	63.7	64.9	63.8	63.2	62.5	59.7	53.6	49.1	49.2

TABLE 9-7 shows the types of cleaning methods employed for U.S. coals for the years 1967-1972.

However, the degree of sulfur reduction attainable via physical cleaning is naturally limited, since only pyritic and sulfate sulfur are removed.

Pyritic sulfur is the mineral pyrite which occurs in coal as discrete particles, although often of microscopic size. It is a heavy mineral which has a specific gravity of about 5.0, compared to coal which has a maximum specific gravity of only 1.7. The pyrite content of most coals can be reduced significantly by utilizing coal preparation methods of size reduction and gravimetric separation.

The U.S. Bureau of Mines performed a comprehensive study (4) of the washability of U. S. coals, involving 455 coal samples from the 6 principal coal producing areas. The standard against which the coal cleaning processes and the washability of the coals was judged was the present federal standard of performance for solid fossil fuel fired steam boilers: Maximum SO_2 emission rate of 1.2 lb/MM Btu heat imput. On this basis the U.S.B.M. concluded:

TABLE 9-7. METHODS OF CLEANING U.S. BITUMINOUS COALS AND LIGNITE*

	Millions of Tons Cleaned					
	1967	1968	1969	1970	1971	1972
Jigs	160	159	155	140	155	128
Tables	50	47	45	44	36	40
Classifiers	4	5	3	4	2	3
Launders	5	4	5	5	5	5
Flotation	8	9	10	10	9	13
Pneumatic	21	17	19	18	15	12
Dense Media:						
Magnetic	65	71	72	77	69	75
Sand	33	27	24	23	18	15
CaCl	3	2	2	2	2	2
Total	349	341	335	323	281	293

*From Item 11, pp. 1-3, in reference list

- "If all the coals were upgraded at a specific gravity of 1.60, the analyses of the clean coal products of the various regions would range on the average from 5.1 to 8.3 percent ash, 0.10 to 1.80 percent pyritic sulfur, 0.56 to 3.59 percent total sulfur, 12,799 to 14,264 Btu per pound and would produce 0.95 to 5.5 pounds of SO_2/MM Btu at Btu recoveries ranging from 91.7 to 97.6 percent. The corresponding SO_2 removal efficiencies required to comply with the current EPA emission regulations of 1.2 pounds SO_2/MM Btu would range from none to 78 percent." The evaluation data are summarized in TABLE 9-8.
- The 455 U.S. coal samples evaluated contained on the average 1.91 percent pyritic sulfur and 3.02 percent total sulfur. Only 14 percent of the raw coal samples could meet the current EPA SO_2 emission standard of 1.2 pounds SO_2/MM Btu with no preparation. If a 50 percent Btu recovery was acceptable, then 32 percent of the samples could be upgraded to meet the standard when crushed to 14-mesh top size.

When these results are contrasted with the most stringent state standard (Wyoming) of 0.2 lbs SO_2 /MM Btu, it is obvious that physical coal cleaning may be considered only as a partial solution for the SO_2 emissions problem; however, as a first step, it will lighten the load on the downstream sulfur removal/recovery systems.

Desulfurization of Coal During Combustion in a Fluidized Bed--
Although fluidized bed combustion (FBC) techniques were developed primarily for the electric power industry so that steam generator systems could utilize high sulfur coals, the principles of FBC may be applied in coal conversion plants to incinerators that incidentally produce steam or to utility boilers whose prime purpose is to generate process steam.

TABLE 9-8. SUMMARY OF COMPOSITE PRODUCT ANALYSES BY REGION FOR CRUSHED AND CLEANED COALS

(Coals crushed to 3/8 inch top size and cleaned at 1.60 specific gravity.)

	Cumulative analyses of float 1.60 product						
	Percent						
	Btu recovery	Ash	Pyritic Sulfur	Total Sulfur	Pounds SO_2/MM Btu(1)	Calorific content, Btu per pound(2)	SO_2 removal efficiency required percent(3)
Northern Appalachian	92.5	8.0	0.85	1.86	2.7	13,766	56
Southern Appalachian	96.1	5.1	.19	.91	1.3	14,197	8
Alabama	96.4	5.8	.49	1.16	1.7	14,264	29
Eastern Midwest	94.9	7.5	1.03	2.74	4.2	13,138	71
Western Midwest	91.7	8.3	1.80	3.59	5.5	13,209	78
Western	97.6	6.3	.10	.56	.9	12,779	None
Total United States	93.8	7.5	.85	2.00	3.0	13,530	60

(1) Based upon the moisture-free Btu value for the float coal and assuming all the sulfur is converted to SO_2. Actual emissions will vary depending upon the as-fired coal moisture content and the amount of sulfur that actually goes out the stack as SO_2.

(2) The calorific content (moisture-free basis) was used to calculate the SO_2 removal efficiency required.

(3) SO_2 removal that must be accomplished in treatment of flue gases to meet the Federal new source performance standard. Values may require adjustment to account for the as-fired coal moisture content and the amount of sulfur that actually goes out the stack as SO_2.

In general, an FBC system will include the following major components:

- The fluidized bed is a mixture of inert bed material (i.e., sand, alumina, coal ash, etc.), coal particles and and SO absorbent, typically limestone or dolomite. The particle size in the bed can vary from -1/4 inch to 100 mesh. The static bed depth can vary between 0.5 and 3 feet with an expanded depth of 1 to 6 feet. The bed is fluidized with a superficial gas velocity between 1 and 15 ft/sec and operates in a temperature range of 1,400-2,000°F.
- Heat transfer surfaces are located both within the fluidized bed and external to the bed. Internal bed surfaces include rows of tubes placed vertically, horizontally or at some intermediate pitch. Membrane type water walls can also be used for internal bed surface. External surfaces include tubes placed directly above the bed and downstream of the combustion zone, such as economizer surface. The temperature profiles in the tubes depend upon the location of the surface and the conditions of the internal fluid.
- Solids handling systems include the transfer of feed materials to the fluidized bed, the removal and/or recycling of bed material, and the return of elutriated solids to the bed. The feed materials injected into the fluidized bed are at ambient temperature and can include coal, limestone, or dolomite, inert bed material and additives. All other systems require handling hot (∿bed temperature) solids.
- Elutriated material recovery. Solids elutriated from the bed include unburned carbon and other bed material. Unburned carbon can be collected and combusted in a carbon burnup cell or simply returned to the primary combustion bed. Other material is usually collected in cyclone type separators and either discarded or returned to the bed with or without further processing. In some cases, baffles are

placed above the fluidized bed to eliminate entrainment of particles in the combustion gases.

The coal-burning fluidized bed combustion boiler is formed by an enclosure usually consisting of waterwalls (abutting boiler tubes in which water and steam flow), or a lined uncooled shell. The pressure in this enclosure is approximately 1 atmosphere. A plate distributes the air flow uniformly over the base of the enclosure. The air then passes at velocities in the range of 1 to 15 ft/sec through a bed of particles at temperatures from 760 to 1,095°C. These particles, whose maximum size is generally 1/4 inch, are comprised of ash or other inert material, both reacted and unreacted sorbent (lime, dolomite) and small quantities (less than 3 percent) of unburned coal or carbon. Coal, sorbent, inert bed material and other additives are fed to the bed by various techniques through the air distributor or the containment shell. From 50 to 60 percent of the heat released in burning the fuel with air is transferred to the water/steam in the tubes surrounding and submerged in the bed.

The air and combustion gases passing through the bed cause considerable agitation. Particles are thrown from the bed into the empty volume above, called the freeboard. Larger, heavier particles fall back into the bed. Smaller, lighter particles are carried out of the enclosure by the gases. Much of the ash formed during the combustion of washed coals is eventually carried from the bed in the combustion gases. Larger ash particles may accumulate in the bed and require continuous removal.

Convective heat transfer surfaces can be located in the path of the combustion products to generate more steam from the sensible heat of these gases. A reasonably high heat transfer coefficient from the combustion gases to the tube surface requires high gas velocities. These velocities can be obtained by narrowing or restricting the gas passage in the convection section or by packing tubes into a broad gas passage with small clearances between tubes.

Ash, fragments of the limestone/dolomite sorbent, and unburned coal char or carbon carried from the fluidized bed enclosure by the combustion gases can be captured by particulate collectors. To obtain high combustion efficiencies, these captured particles must be recycled to the boiler bed or to a separate fluidized bed combustor (a carbon burnup cell) where burning of the char is completed. A secondary collector can collect particles not collected in the primary collector for disposal.

Additional heat can be extracted from the combustion gases in the heat recovery section, which may be an economizer (that preheats boiler feedwater) and/or an air preheater. Final cleanup of the combustion gases is carried out in a secondary particulate removal system.

FBC can be operated in several ways to effect the desired solid-gas reactions:

- Downflow, where limestone and fuel are fed into a precalcined bed of lime. Gas velocities are usually low so that little lime, but most of the coal ash, is elutriated. Solids, consisting of $CaSO_4$, CaO, and some ash, are removed through an overflow pipe, thus maintaining a constant bed level. The cooling surface can be immersed directly in the bed to maintain the bed temperature at the desired level.

- Upflow, where the bed consists of coarse inert material. Gas velocities are higher and feed materials are finer so that both spent lime and ash are carried overhead. Because the solids retention time is short, this mode of operation is suitable if the gas-solids reactions are rapid.

- Multi-Solids Combustor, under development by Battelle Memorial Institute, uses a high specific gravity material for the base bed with a recirculating entrained bed of fine particles above it. Coal and limestone are fed into the dense bed, SO_2 is absorbed there and the released heat maintains the temperature of the entrained bed in which the steam tubes are located. Space velocities up to 34 ft/sec have been reported. Satisfactory SO_2 removal with Ca:S ratios as low as 1.4 is claimed.

- Cyclonic, patented by Babcock & Wilcox, is designed with a high velocity center tube through which all combustion gases pass and an annular fluidized bed of lime. The configuration allows feed to enter at the top of the annular bed then the char produced enters the bottom of the central tube to supply heat to the solids while the combustion air maintains a high velocity vortex in and at the top of the tube. High unit capacity is claimed.

Operating temperatures of 760 to 870°C are predicted according to limestone calcination theory and have been proved in practice: temperatures above 760°C are required to promote rapid combustion and for reasonable SO_2 reaction rates, while above 870°C sorbent activity appears to decrease and corrosive alkali metal salts begin to volatilize.

Best operation is obtained with excess air ranging from 5 to 15 percent. Above 15 percent, NO_x formation becomes a problem and excessive heat is lost in the stack gases. Below 5 percent leads to emissions of combustibles and freeboard burning.

When coal is burned with excess air in the pressure of calcined limestone or dolomite the sulfur in the coal is oxidized to SO_2 and reacts with limestone to form calcium sulfate or with dolomite to form a mixture of calcium and magnesium sulfates. Although thermodynamic studies indicate that sorption of SO_2 or H_2S by limestone is possible and should proceed nearly to completion, experimental results indicate that calculated equilibrium values are not approached and that the reaction kinetics appear to be controlling. Careful design based on both theoretical and actual considerations is required for satisfactory performance.

Reference 7 discusses the design factors, their application and results of their application on pp.3-1-3-9 and 6-1-6-2.

Flue Gas Desulfurization--
If sulfur in fuels fed to combustion equipment is not removed before or during combustion by methods discussed previously, then flue gas desulfurization (FGD) must be employed to meet stringent environmental standards.

The average U.S. coal contains 3.02 percent total sulfur and has a higher heating value of 12,574 Btu/lb (4,p.2). If this fuel is burned without controls, an SO_2 emission of 4.8 lb/MM Btu results. To meet the most stringent standard of 0.2 lb/MM Btu, an FGD efficiency of about 96 percent would be required. Obviously, if the coal contains less sulfur a lower SO_2 removal efficiency is required, and conversely.

The tar produced from a gasification facility processing average U.S. coal is likely to contain about 1.81 percent sulfur (60 percent of 3.02 percent). The uncontrolled SO_2 emission would be about 2.2 lb/MM Btu. Therefore, to meet a standard of 0.34 lb/MM Btu, an FGD efficiency of about 85 percent is required.

More than 100 FGD processes in various stages of development are reported in the literature (9,p.67). A Kellogg study addressed such processes and recommended 4 for further study (8). These were:

- USBM Citrate
- Wellman-Lord
- Chiyoda Thoroughbred 101
- Pullman Kellogg Weir

The Federal Power Commission, reporting on 7 commercial FGD processes, listed the process charactistics (9,p.40) as in TABLE 9-9.

In a later section of the report they describe 15 advanced FGD processes which are estimated to be commercially available from 1976 to 1988. These are shown in TABLE 9-10.

There were about 45 FGD units installed at 11,000 MW of generating capacity as of the end of 1977. Projections indicate that about 70 FGD units, installed on 26,000 MW of generating capacity, will be in operation by 1981 (9,p.50). In 1975, the weighted average capital cost of the operating FGD systems was $90/KW and the operating cost was 3.1 mills/KwH (9,p.24). The costs were somewhat higher for retrofitted systems as opposed to new systems and for systems treating high sulfur coal as opposed to those treating low sulfur coal.

FGD processes are normally subdivided into two types:

- Regenerable, where the product sulfur removed is in marketable form such as elemental sulfur (S), sulfuric acid (H_2SO_4), or concentrated sulfur dioxide (SO_2).

- Non-regenerable, where the product sulfur removed is in waste form (normally a sludge containing $CaSO_4$, $CaSO_3$, unreacted $CaCO_3$, fly ash, and water).

TABLE 9-9. CHARACTERISTICS OF COMMERCIAL FGD PROCESSES

Process	Primary Removal Agent	Regenerable?	Form of Principal Sulfur Product	Operational Mode
Wet Lime/Limestone	Lime or Limestone	No	$CaSO_3$ (Throwaway)	Wet
Alkaline Fly Ash	Alkaline Fly Ash	No	$CaSO_3$ (Throwaway)	Wet
Sodium Carbonate	Sodium Carbonate	No	Na_2SO_3(Throwaway)	Wet
Double Alkali	Sodium Hydroxide	Yes	$CaSO_4$ (Throwaway)	Wet
Dilute Sulfuric Acid/Gypsum	Sulfuric Acid	Yes	$CaSO_4$ (Throwaway)	Wet
Magnesium Oxide	Magnesium Oxide	Yes	SO_2	Wet
Wellman-Lord	Sodium Sulfite	Yes	SO_2	Wet

TABLE 9-10. CHARACTERISTICS OF ADVANCED FGD PROCESSES

Process	Primary Removal Agent	Regenerable?	Form of Primary Sulfur Product	Operational Mode
Agglomerating Cone	Phosphate Rock Slurry	No	Fertilizer	Wet
Allied/Wellman-Lord	Sodium Sulfite	Yes	Sulfur	Dry
Ammonia Scrubbing	Ammonium Solution	Yes	Sulfur	Wet
Basic Aluminum Sulfate - Gypsum	Aluminum Sulfate	Yes	Gypsum	Wet
Catalytic Oxidation	Vanadium Pentoxide	Yes	H_2SO_4	Dry
Citrate	Citric Acid	Yes	Sulfur	Wet
Copper Oxide (Shell)	Copper Oxide	Yes	SO_2	Dry
Dry Adsorption	Activated Carbon	Yes	Sulfur	Dry
Electrolytic Regeneration (Stone & Webster/Ionics)	Caustic Soda	Yes	H_2SO_4	Wet
Manganese Oxide	Manganese Oxide	Yes	SO_2	Dry
Aqueous Carbonate	Liquid Carbonate	Yes	H_2S	Wet
Nahcolite Injection	Nahcolite Ore	No	Na_2SO_4	Dry
Organic Absorbent	Glyoxylic Acid	Yes	SO_2	Wet
Potassium Thiosulfate	Sulfates, Sulfites	Yes	Sulfur	Wet
Phosphate	Phosphate Buffer	Yes	Sulfur	Wet

This study will be confined to the description of several of the more promising of the candidate systems.

Citrate Process--Research sponsored by the U.S. Bureau of Mines has developed a process where the flue gas is first cooled and scrubbed with an ash slurry to remove solid particles and sulfur trioxide. The gas then enters an absorber where over 95 percent of the SO_2 contained in the treated gas is removed by contact with a citric acid/sodium citrate/sodium thiosulfate solution. The clean gas is reheated and exhausted to the atmosphere.

The SO_2 rich solution flows by gravity to a stirred reactor vessel and is reacted with H_2S, which is available from the acid gas removal system, to precipitate sulfur. The resultant slurry is separated by flotation as a 10 to 15 percent solids product, leaving clear regenerated citrate for recycle to the absorption tower.

The sulfur flotation product is filtered and the solids are heated to melt the sulfur at about 138°C. The hot liquid passes into a settler tank from which molten sulfur is tapped. The citrate solution is withdrawn from the top of the settler for reuse (8).

Advantages of the citrate process include:

- Removal efficiencies on the order of 95 to 98 percent have been obtained on pilot scale operations.
- Precipitation of sulfur compound takes place outside of the boiler; hence, there is no plugging or scaling.
- The system has a high capacity for short-term SO_2 overloads.

- The reactions in the reactor tank occur rapidly, resulting in a fast-settling sludge.

- The low liquid-to-gas ratio leads to a low pressure drop across the absorber. Energy requirements are thus minimized.

- The product is elemental sulfur.

The process has some disadvantages and possible problem areas:

- The process of separation of sulfur from solution in a decanter may be difficult.

- Regeneration by production of H_2S may prove to be costly due to expenses associated with the gas feed requirements. However, when H_2S is available, this is no longer a factor.

- Lack of commercial experience may be the main negative factor (9,p.72).

Although no commercial plants are presently in operation, one is being installed at the St. Joe Minerals Company's G.F. Weaton Power Station at Monaca, Pennsylvania. This unit will treat 156,000 SCFM (234,000 ACFM) of flue gas from a 60 MW coal fired boiler. Completion of construction is scheduled for the fall of 1978. A one year demonstration program will follow.

Capital and operating costs for an FGD unit handling flue gas from generation of 1000 MW fed with 2.5 percent sulfur coal are estimated to be \$73/KW and 1.97 mills/KWH or \$5.49/ton of coal (1977). These figures include the H_2S generation unit. If H_2S generation is not needed, the capital cost drops to about \$67/KW and the operating cost is also reduced (10,pp. 1-4,25-27).

Allied/Wellman-Lord Process--In this process the flue gas is first scrubbed by a recirculating ash slurry to remove fly ash. The concentration of the circulating stream is controlled by a purge stream which is directed to a disposal pond.

The cool gas then enters an absorber where it is contacted with a sodium sulfite solution which absorbs SO_2 to form sodium bisulfite. A side stream of this solution is purged for sulfate removal by crystallization. The clean flue gas is discharged to a stack after reheating.

The absorber liquor is heated in an evaporator-crystallizer, where SO_2 and water vapor are released and sodium sulfite crystals are removed as a slurry. The water in the overhead vapors is condensed and used to redissolve the crystals for recycle to the absorber. Prior to redissolving, the slurry is thickened and the clear solution is returned to the evaporator. A small purge is withdrawn to control the thiosulfate concentration. The SO_2 gas is compressed and can be sent to a Claus plant where it reacts with H_2S available from the complex to recover sulfur.

Advantages of the process are:

- Elemental sulfur is recovered in the process, allowing easy handling and yielding a potentially marketable product.
- The operation can quickly adjust to flue gases of any SO_2 concentration.
- The process has good commercial experience with 10 units recovering SO_2 in operation. Another 18 units are in various design stages.

- SO_2 removal efficiencies of greater than 95 percent are possible.

Possible disadvantages and problems are:

- The use of methane as a reducing gas can lead to significantly high materials costs. It is not required when H_2S is available.

- Capital cost for this process appears somewhat higher than for others. A Pullman Kellogg study indicated that the capital investment is about 33 percent higher than the citrate process (8) (9,pp.67,70).

Ammonia Scrubbing Process--In this process ammonia compounds, e.g., ammonium bisulfite, possessing very high SO_2 absorption rates, are used as the absorbent. The sulfites and sulfates thus produced can be regenerated by chemical reduction. Elemental sulfur is recovered by the Claus process by conversion of recovered SO_2 with hydrogen sulfide recovered from the complex (9,p.70).

Process advantages are:

- The very high SO_2 absorption rate for ammonia permits a low liquid-to-gas ratio.

- Extensive regeneration of the spent scrubber solution is possible due to the volatile nature of ammonia.

Possible disadvantages and operational problems are:

- An environmentally hazardous plume tends to form if process control is not rigidly maintained. Ammonium sulfate has been found to be the primary constituent of the exhaust gas.

- In practice it has sometimes been difficult to separate the fly ash and the ammonium sulfite.

Shell FGD Process--In this process, the flue gas is first passed through precipitators, of a type determined by the particulate size distribution, for solids removal.

Flue gas then passes over a static, packed bed of copper oxide acceptor plates at about 400°C. SO_2 and oxygen in the flue gas react to form copper sulfate. When one bed is spent, it is regenerated with hydrogen, also at 400°C. The reactions are summarized as follows:

$$CuO + 1/2\ O_2 + SO_2 \longrightarrow CuSO_4$$

$$CuSO_4 + 2H_2 \longrightarrow Cu + SO_2 + 2H_2O$$

$$Cu + 1/2\ O_2 \longrightarrow CuO$$

The regeneration off-gas passes through a quench-absorber stripper section which reduces the temperature of the gases by waste heat recovery and delivers a concentrated SO_2 stream for further processing.

If recovery of elemental sulfur is required, then H_2S from the facilities is used for the Claus reaction (8) (9,p.73).

Advantages of the process are:

- Since this is a dry process, there is no handling of wet materials and water input requirements are minimal.
- Acceptance and regeneration occur at approximately the same temperature, obviating any heating or cooling of the absorption beds.
- The flue gas passes over the surface of the acceptor material rather than through it. This arrangement prevents plugging of the acceptor bed.
- By alternating the acceptor and regeneration units, continuous processing can be maintained.
- Operating costs are minimal due to absence of reheating and to low water requirements.
- The process has the potential for expansion to SO_x and NO_x removal by ammonia injection into the acceptor bed. The CuO will act as a catalyst for NO_x absorption.
- There is no disposal of spent sludge.

Process disadvantages and possible operating problems include:

- Equipment and installation costs are high.
- A hydrogen source is needed for regeneration. Operating costs may be high due to the large hydrogen requirements.
- Recovery is in the form of gaseous SO_2 , requiring further processing for marketing.

- Damper operation becomes important when switching operational modes at high temperature.

- The stripper requires larger quantities of steam resulting in high energy input requirements. The acidic wastewater must be neutralized prior to discharge.

- The hydrogen feed may not be adjustable to fluctuations in SO_2 input concentrations.

Pullman Kellogg Weir Scrubber--Pullman Kellogg offers a multistage, countercurrent, limestone scrubbing process with a low pressure drop, high SO_2 conversion and high limestone utilization. In the first step, hot gases are quenched to saturation temperature and passed along a series of horizontal scrubbers where a limestone slurry is sprayed vertically into the contact chamber. The gases are reheated to discharge temperature and enter the stack for discharge to the atmosphere.

The limestone slurry, containing catalysts, is collected in a slurry tank and is recirculated to the scrubber. Net make of sulfite-sulfate slurry is oxidized by air or oxygen and decanted. Thickened slurry is filtered on a rotary vacuum filter and the cake is either directed to disposal or further processed for marketing as commercial gypsum. Clear liquid is sent to limestone slurry makeup. Catalyst, usually magnesium salts or oxide, must be added to compensate for losses due to retention of some liquid in the washed filter cake (8) (9,p.43).

The process has the following advantages:

- Preliminary economics indicate that this process has one of the lowest capital outlays of all the processes being considered.

- The basic process is fairly simple. Very few process steps are included.

- Reserves of absorbent materials are fairly abundant in the United States.

- SO_2 removal efficiencies are generally high, on the order of 95 percent.

- The two-stage treatment of flue gases allows for the simultaneous removal of SO_2 and particulates.

- The process is the method most commonly used by utilities for SO_2 control, exclusive of low sulfur fuel. Commercial installations have been operating for more than four years. Therefore, the lime/limestone process is the most fully characterized of the existing FGD processes. Operational experience has led to a greater understanding of basic principles.

- Successful performance, particularly in terms of SO_2 removal, on coal-fired systems has been demonstrated.

- The process is not adversely affected by fly ash in the system.

Disadvantages and possible process problems are:

- Large quantities of waste require disposal in an environmentally acceptable manner.

- If not designed carefully or operated attentively, lime/limestone systems have a tendency towards chemical scaling, plugging, and erosion. These problems can frequently halt operation of the system.

- There appears to be a deficiency in the understanding of the factors that cause or prevent the oxidation of sulfite to sulfate, thus enhancing or inhibiting serious scaling. Excess air, pH, fly ash, residence time in the reaction tank, and the presence of NO_2 in the flue gas are suspected to be contributing factors.

- Very high liquid-to-gas ratios are required in the scrubber.

- In the wet lime process, the sludge has poor settling properties due to the high sulfite content.

Chiyoda Process--The Chiyoda Thoroughbred 101 Process requires a front-end water scrubber to eliminate solids carryover in the flue gas. The SO_2 in the gas is absorbed in 2 to 5 percent sulfuric acid where it is oxidized to SO_3 and H_2SO_4 by air in the presence of a ferric sulfate catalyst. The gas is reheated for release through the stack.

The net make of acid is neutralized with lime or limestone, producing gypsum which is separated by filtration. The mother liquid is returned to the scrubbing step. The gypsum is of high grade, suitable for use in wallboards or cement (8) (9,p.46).

Advantages of the process are:

- The process flow and plant structure are uncomplicated. Capital cost is reasonably low.

- If gypsum can be recovered as a marketable product, the need to dispose of waste products is obviated. If it cannot be recovered, the form of the waste product is relatively easy to handle.

- High SO_2 removal efficiencies, on the order of 97 percent, have been reported.
- Scaling problems have proved to be minimal, relative to those encountered with the lime/limestone process.
- No plugging problems have been encountered, since the circulating fluid is highly acidic.
- The operational reliability of this process has been relatively high.
- The gypsum produced is of good quality.
- The process has good commercial experience.
- The process is not adversely affected by fly ash in the system.

Disadvantages and possible operational problems are:

- Since the process involves handling sulfuric acid solutions, special corrosion-resistant materials are required.
- There is a possibility of problems resulting from poisoning of the ferric sulfate catalyst.
- The marketability of gypsum in the United States is highly questionable.
- The process design requires large pumps and fairly large scrubbers and oxidizers.
- A high liquid-to-gas ratio is required in the scrubber.

- The recovered gypsum is dewatered in a centrifuge which requires high maintenance.

- Handling difficulties arise from the large volumes of sulfuric acid passing through the system.

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Processes for Control of Hydrogen Sulfide

The Claus Process--

The Claus Process is a commercial process for production of sulfur from gas streams that contain H_2S. The system is very flexible with regard to the H_2S content of the feed gas: H_2S concentrations as low as 7 percent (1, p.124) and as high as essentially 100 percent can be sucessfully treated. Conversion of H_2S to sulfur can be as high as 95 to 97.5 percent for three-stage plants treating rich feed streams (6, p.137), while the conversion will fall off to about 90 to 94 percent for lean feed streams.

The process essentially consists of an oxidation step, where one-third of the H_2S is oxidized with air to SO_2, followed by several catalytic reaction steps where the rest of the H_2S reacts to form sulfur. Overall, the reaction can be represented by the equation:

$$3H_2S + 1.5\ O_2 \longrightarrow (3/x)S_x + 3H_2O$$

The presence of hydrocarbons in the feed gas can cause a deterioration of the plant operation in that:

- Conversion of H_2S to sulfur is reduced
- Combustion air requirements are increased
- Gas volume flowing through the plant can increase substantially

  For example, a feed containing 5 percent methane will increase air requirements by 35 percent and will increase gas flow by 27 percent (1, p.120)
- COS and CS_2 may be formed from the hydrocarbons (1, p.121)

Normally it is recommended that the hydrocarbon content of the feed be reduced to a low level prior to the Claus operation.

The presence of ammonia in the feed gas presents a potential problem. For a conventional Claus plant, 2 to 3 percent ammonia in the feed gas is a practical limit. However, when the feed gas contains 30 percent or more of CO_2, the ammonia content should be maintained below 0.05 to 0.10 percent (1, p.121). Ammonia can be removed from the feed gas by scrubbing with water.

For these various reasons, a Claus plant alone cannot be expected to reduce H_2S in process offgases to a level low enough to meet environmental standards. Developments in technology for control of H_2S have been directed toward treatment of Claus offgas and have been shown to be successful. Combination of offgas treatment processes with the basic Claus process, therefore, results in systems that will, in most cases, yield final offgas streams that can be vented to atmosphere.

Combined Claus and Beavon Processes--

This design combines a Claus unit for bulk sulfur conversion and recovery with a Beavon unit for conversion of the remaining sulfur compounds present in the Claus tail gas to elemental sulfur.

The Beavon process is a tail gas cleanup process based on the catalytic conversion of sulfur species to H_2S by hydrogenation and hydrolysis followed by conversion of H_2S to elemental sulfur by the Stretford process. The process is commercial and is licensed by The Ralph M. Parsons Company. There are about thirty Beavon installations in operation (4,p.4). It is highly successful in reducing sulfur emissions to the atmosphere. Effluent from the absorber is guaranteed to contain less than 100 ppm (v) total sulfur of which less than 10 ppm is H_2S.

Combined Claus and SCOT Processes--

This combination links a Claus plant to the SCOT (Shell Claus Off-Gas Treating) process for tail gas treatment. The SCOT process treats the gas by a catalytic conversion step which converts various sulfur species (COS, CS_2, SO_2, S, etc.) to H_2S via hydrogenation and hydrolysis reactions. Fuel gas (light hydrocarbon) and air are needed to produce the reducing gas for these reactions, as in the Beavon process. The H_2S is then absorbed into an alkanolamine solution in a scrubbing system by countercurrent contact with the solvent. The overhead gas stream is released to atmosphere.

Rich solvent is pumped through a heat exchanger to the stripper where H_2S is released overhead in a fairly concentrated stream that is recycled to the Claus plant. Steam supplies heat to the stripper reboiler and cooling water is normally used in the overhead condenser.

The process is commercial with a number of installations in operation. It is licensed by Shell Development Company. A high sulfur cleanup efficiency is supposedly achievable by the process. It is claimed that the absorber overhead gas contains less than 100 ppm (v) total sulfur species with less than 10 ppm (v) H_2S. Reported data, however, indicate that the total sulfur content of this stream may be as high as 300 to 350 ppm (v) (7).

Combined Claus and ARCO Processes--

This design combines a Claus plant with an ARCO (Atlantic Richfield Company) process for tail gas treatment. No performance data are available on the process, but information from Parsons (2) indicates that the concept is essentially the same as the Claus/SCOT combination.

Combined Claus and Incineration Processes--
This system simply combines a Claus unit with an incineration step to oxidize the tail gas sulfur species to SO_2. A flue gas desulfurization (FGD) process is required following incineration. This concept has the advantage of removal of hydrocarbons (by oxidation) which may otherwise be present in the tail gases. It is, of course, commercial.

Combined Claus and Lucas Processes--
This system combines a Claus plant with a Lucas plant to treat the Claus tail gas. In the Lucas process, all sulfur species present in the Claus tail gas are incinerated to SO_2. The SO_2 is recovered in concentrated form and returned to the Claus plant.

The Lucas process consists of three basic stages. In the first (incineration), tail gas is burned with air and fuel to convert most of the sulfur compounds to SO_2. The second stage consists of a hot coke treatment of the incinerator off gas where the following reactions take place:

$$O_2 + C \rightarrow CO_2$$
$$2SO_3 + C \rightarrow 2SO_2 + CO_2$$
$$H_2S + 1.5O_2 \rightarrow SO_2 + H_2O$$
$$2NO + C \rightarrow N_2 + CO_2$$

The third stage is the removal and concentration of SO_2 by an absorption/regeneration process using aqueous alkali phosphate solution. Sulfur dioxide is absorbed as in the following reaction:

$$Na_2HPO_4 + H_2O + SO_2 \rightarrow NaH_2PO_4 + NaHSO_3$$

In the regenerator the reaction is reversed and SO_2 is released for recycle to the Claus unit (8, p.110).

The process has been tested on a semi-commercial scale in a German refinery. To our knowledge it has not yet been applied on a fully commercial scale. The absorber off-gas supposedly contains practically no H_2S, less than 200 ppm SO_2 and less than 150 ppm COS and CS_2. From the standpoint of pollution control, this process does not appear to compare favorably with those previously discussed.

The Stretford Process for Primary Sulfur Recovery--
As an alternate to primary sulfur production by Claus processing, the Stretford process may be employed in one of the following manners:

- Pressure Stretford process for removal of H_2S from the main process gas stream and subsequent sulfur production.

- Low pressure Stretford process for sulfur production from an H_2S-rich stream from an acid gas removal (gas purification) system such as Selexol, Rectisol, MEA, hot carbonate, etc.

The Stretford process is described in detail in "Integrated Schemes for Emission Control," later in this section of the report. When the Stretford process is employed, some means must be employed to convert organic sulfur compounds (COS, CS_2, mercaptans,and thiophenes) to H_2S upstream of the Stretford unit, otherwise they would pass through the Stretford absorber to the atmosphere. A number of catalytic conversion processes are available for this purpose. Some of these are:

- Holmes-Maxted (commercial)

- Carpenter-Evans (commercial but not widely accepted)

- British Gas Council (proposed)

In these processes the gas is heated and passed through a catalyst bed where organic sulfur compounds are catalytically converted to H_2S by hydrogenation and hydrolysis. The heated gas is then suitable for feed to the Stretford process. A simplified flow scheme and the reactions taking place are shown in Figure 9-34.

When a hot potassium carbonate process is used for H_2S removal from process gas, a considerable degree of hydrolysis of COS occurs:

$$COS + H_2O \rightarrow H_2S + CO_2$$

In this case the H_2S-rich gas may be suitable for use in the Stretford process without a catalytic conversion process.

The Stretford process itself is commercial with 31 plants in operation (5). A later article (1978) reports 36 plants in operation. Conversion efficiencies for H_2S are reported to be over 99% and H_2S concentrations in the absorber off-gas are reported to be between 5 and 8 ppm (v).

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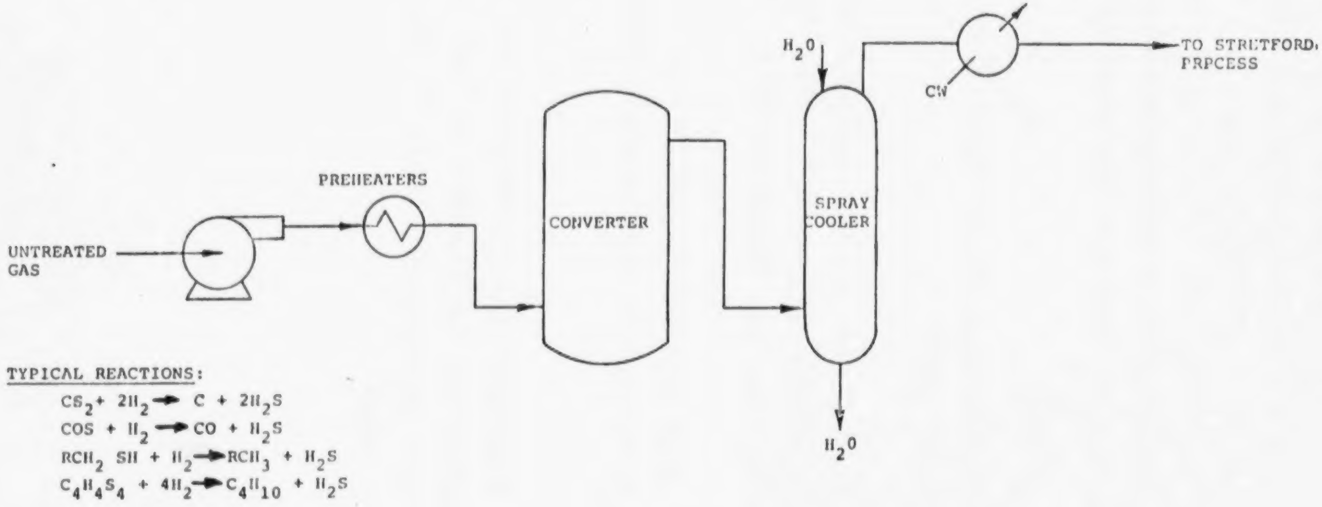
UNTREATED GAS
PREHEATERS
CONVERTER
H_2O
SPRAY COOLER
H_2O
CW
TO STRITFORD PRPCESS
TYPICAL REACTIONS:
$CS_2 + 2H_2 \rightarrow C + 2H_2S$
$COS + H_2 \rightarrow CO + H_2S$
$RCH_2SH + H_2 \rightarrow RCH_3 + H_2S$
$C_4H_4S_4 + 4H_2 \rightarrow C_4H_{10} + H_2S$

Figure 9-34. Typical schematic and reactions for catalytic conversion processes.*
*From Item 9, p. III-31 in reference list

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Techniques for Control of Particulates

Particulates are chemical elements, compounds or mixtures of them in either solid or condensed liquid droplet form. They are usually described in terms of the physical characteristics which affect the mechanisms for their separation. The most important of these are physical size and density. Figure 9-35 is a summary of particulate sources, their size ranges and the equipment usually used to remove them from gas streams.

The size, density, and nature of particulate matter directly affect equipment selection and design, as shown in Figure 9-36.

The types of particulate collectors may be arranged in order of increasing particulate collection efficiency, complexity and cost as:

- Cyclone collectors
- Wet scrubbers
- Fabric collectors
- Electrostatic precipitators

Cyclones--

The most widely used type of particulate control device is the cyclone collector. An example is shown in Figure 9-37. Particulate-laden gas enters tangentially at the top of the cyclone body and its liniar velocity is translated into angular acceleration. Because the density of the particulates is greater than the density of the gas, centrifigal force moves them rapidly to the cyclone wall where they collect. The force of gravity encourages movement of the collected particulates downward to the discharge at the cyclone bottom.

Of the many parameters in the design of a cyclone for a specific application, pressure drop and particle collection efficiency are

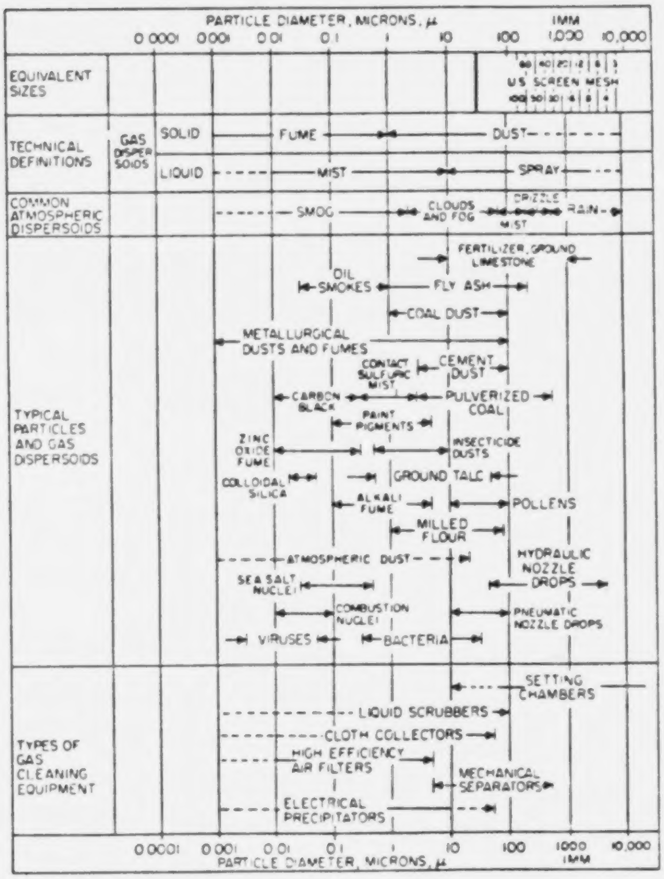

Figure 9-35. Characteristics of particles and particle dispersoids. (Adapted from Stanford Research Institute*

*From Item 1, p. 5-3 in reference list

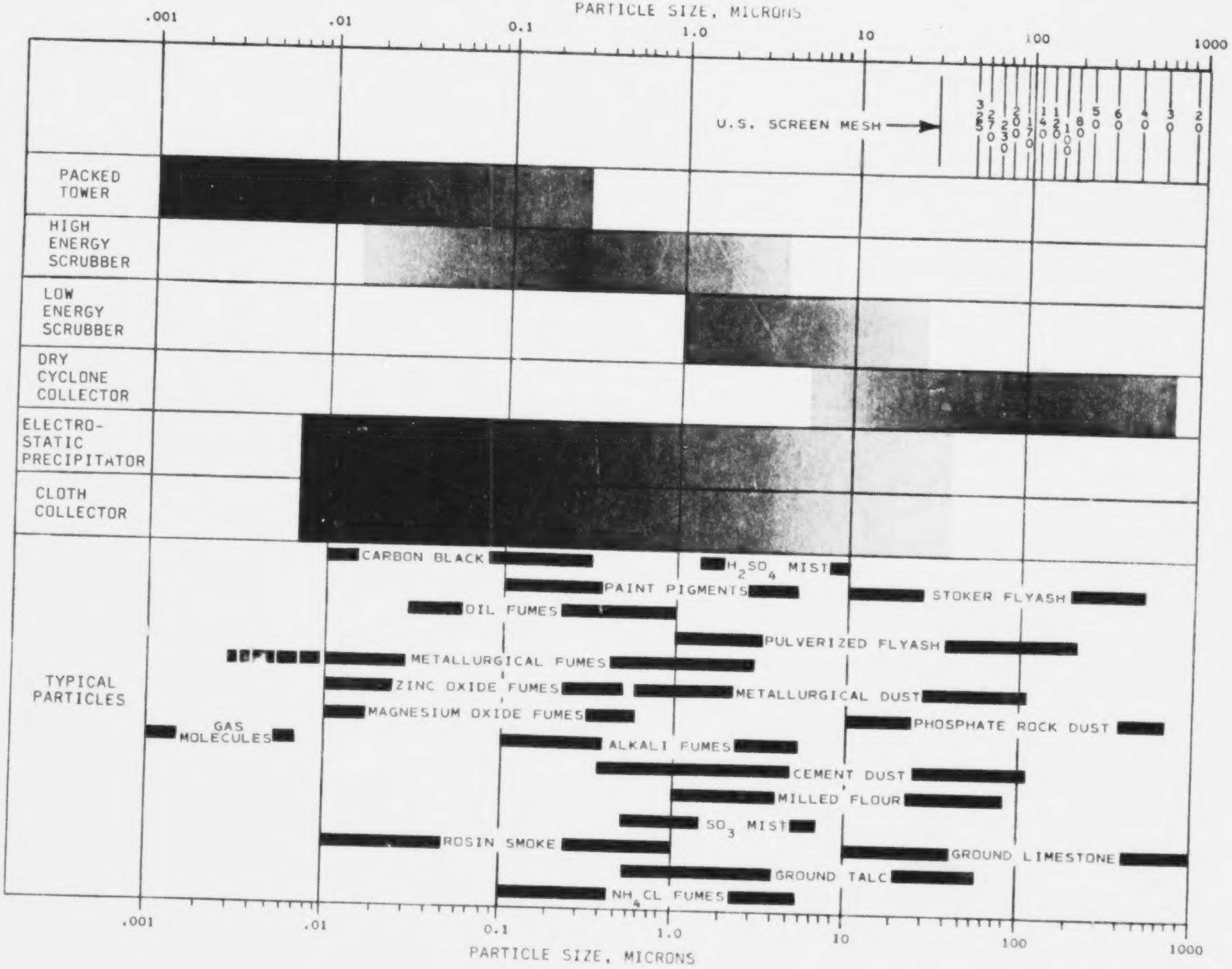

Figure 9-36.- Types of collectors for various constituents.

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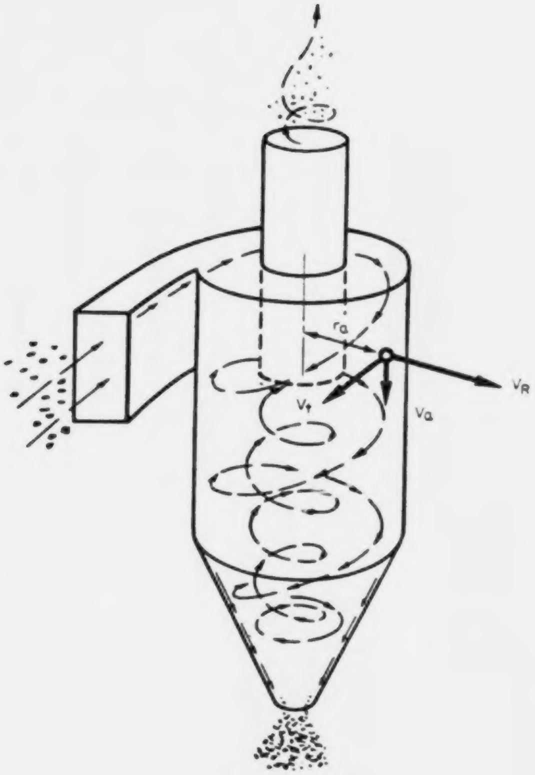

Figure 9-37. Operating principle of a dry inertial cyclone collector.*

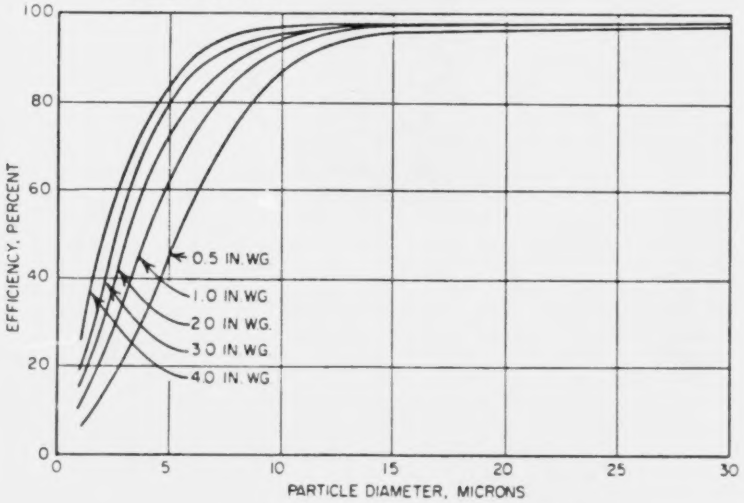

Figure 9-38. Typical fractional performance curves for a multitube mechanical collector.*

*From Item 1, pp. 5-7, 5-9 in reference list the most important, pressure drop because of its effect on duct sizing, fan sizing and fan horsepower, and efficiency because of its relationship to pressure drop and cyclone configuration. Both of these parameters become particularly important when large volumes of gas must be cleaned and have been the incentive for development of cyclone types that depart considerably from the classic configuration of Figure 9-37. One such design subdivides the gas stream among several small cyclones to maintain separation efficiency without excessive pressure drop. The performance of such an array is shown in Figure 9-38.

Collection efficiency as depicted in Figure 9-38 is typical for cyclone collectors in that efficiencies drop rapidly for particles below about 10 microns in diameter. Cyclones are most commonly used in these applications:

- o When particulates are mainly in the coarser size ranges
- o When particulate concentrations are fairly high, e.g., above 3 grains per scf
- o When high collection efficiency is not critical
- o When they can serve as pre-collectors in conjunction with other types of collectors that are more efficient in removing fine particulates

Cyclones have the lowest capital cost of the four general types of particulate collectors. Costs of \$0.08 to 0.10/ACFM* for units in the capacity range of 100,000 ACFM are reported. Special custom designed units may cost as much as \$0.35/ACFM. Installation will usually add about 25 percent to the cost. These figures are based on 1971 data (1, p.5-10).

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*ACFM = Actual cubic feet per minute

Wet Scrubbers--

In particulate collection, wet scrubbers are next in terms of relative complexity and initial cost. There is a great variety of individual types and configurations. The scrubbers operate on the basic principle of confronting the particulates with impaction targets which can be either wetted surfaces or, most frequently, individual droplets. The efficiency of a wet scrubber is a function of several variables. It will be higher when particle diameter, particle density, and relative velocity between the particle and the target droplet are high and when the gas viscosity and target droplet size are low. Thus, to obtain efficient particulate removal from a given stream, it is advantageous to employ a high liquid-to-gas ratio and to produce a high degree of atomization of the scrubbing liquid.

The efficiency of collection of particulates of a given particle diameter and density is related to the energy consumption of the scrubber system, as measured by the pressure drop across the system, and is also related to the manner in which the energy consumption is utilized (1, p.11-11):

	Collection Efficiency at		Pressure Drop,
	5 Microns	2 Microns	cm Water
Spray Tower	94%	81%	8
Orifice Impingement	97	94	8
Self-Induced Spray	93	78	15
Venturi	99.6	97	60 to 75

The orifice impingement scrubber maintains its high collection efficiency with low energy consumption as particle size decreases and is roughly comparable in efficiency to the venturi in efficiency at particle sizes below 1 micron, but with only 10 to 12 percent of the venturi's energy consumption. The efficiency of the spray tower is roughly comparable to that of the self-induced spray but with half the energy consumption. Choice of collector therefore must take into account the particle size distribution of the particulates, the required collection efficiency and the overall energy consumption.

High energy wet scrubbers (venturi scrubbers) are normally used where:

- Fine particles must be removed at high efficiency
- Cooling is desired and moisture addition is not objectionable
- Gaseous contaminants as well as particulates are involved
- Volumes are not extremely high (because of the relatively higher operating cost per ACFM)
- Relatively high pressure drop is tolerable
- Contamination of the scrubbing liquid with materials removed from the gas poses no problem

Initial cost (1971) of wet scrubbers sized for about 100,000 ACFM ranges from \$0.25 to 0.35/ACFM in carbon steel and about \$0.65/ACFM in alloy steel with cost of erection adding about 25 percent (1, p.5-13).

Fabric Filters--

Various configurations of fabric filters are used in which porous fabrics remove particulate from gas streams by allowing clean gas to pass though while preventing the passage of the particles. The buildup of particulate matter on the filter medium aids in the particulate removal process, as shown in Figure 9-39.

Fabric filters are usually arranged as a number of cylindrical tubes, or bags, suspended in an enclosure. Design variations include arrangements for gas passage from the outside or from

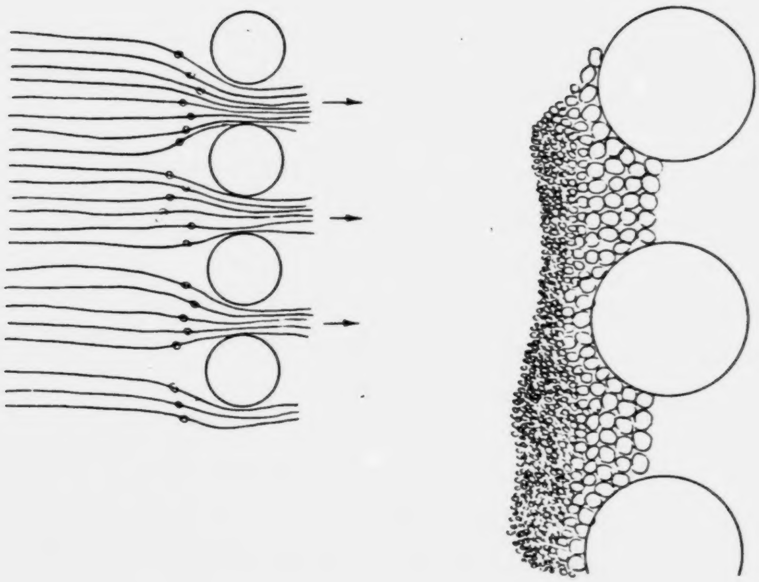

Figure 9-39. Operating principles of a surface type fabric filter. Initial condition at left; filtering effect of "cake" indicated at right.*

From Item 1, p. 5-13 in reference list

the inside of the bags and arrangements for removal of collected dust from the surfaces of the bags. These two principal variations are frequently related: bags cleaned by mechanical shaking and by high pressure air jets inside the bags usually collect dust on the outside of the bags, while bags cleaned by pressurizing a section of the bag enclosure or by high pressure air jets outside the bags collect dust on the inside of the bags. Overall, there is little advantage of one general type over the other, with choice finally being influenced by such considerations as rate of bag wear, ease of replacement, auxiliary power required, etc.

Very high efficiencies can be attained with fabric filters (99+ percent). The principal design considerations are:

- Superficial velocity (air to cloth ratio)
- Fabric resistance coefficient (permeability)
- Cake resistance coefficient
- Weight of cake per unit area
- Gas viscosity
- Means of filter cleaning (mechanical shaking, collapsing, reverse flow, shock wave, pressure pulse, etc.)

Fabric filters are normally employed where:

- High efficiencies are desired
- Operation is above the gas dew point
- Temperatures are moderate
- Valuable material is to be collected dry
- Water availability and disposal is a problem

Initial cost of these units (1971) ranges from $0.50 to 1.20 per ACFM, depending on the filter medium. Erected cost is reported to be about 30 percent of the equipment cost.

Electrostatic Precipitators--

In the electrostatic precipitator (ESP) particles are charged in an electric field, collected on the passive collecting electrode and removed by mechanical means to hoppers located below the collectors. The efficiency of collection is dependent primarily on the total surface area of collecting electrode per unit volume of gas treated and the precipitation rate parameter. The latter is directly proportional to field strength and particle diameter and inversely proportional to gas viscosity. The factors which affect field strength and ultimately the collection efficiency are the gas density and the particle bulk resistivity. High gas density and low particle resistivity result in high efficiency, all other factors being equal. The particle resistivity can be controlled by selection of the proper operating temperature or by use of conditioning agents. The resistivity of fly ash as a function of temperature and coal sulfur content is shown in Figure 9-40.

ESP's are most commonly used when:

- Very high efficiencies are required for fine materials
- Volumes of gas are very large
- Water availability and disposal are problems
- Valuable dry material is to be recovered

The purchase price (1971) of an ESP in the 100,000 ACFM capacity range is about $0.80/ACFM while that of one ten times as large is about $0.40/ACFM. Erection cost adds about 70 percent (1, p.5-20).

Sources and Control of Particulate Emissions--

The major particulate emissions from a coal conversion facility are expected to come from the following sources:

- Coal storage and reclaiming facilities, estimated to be

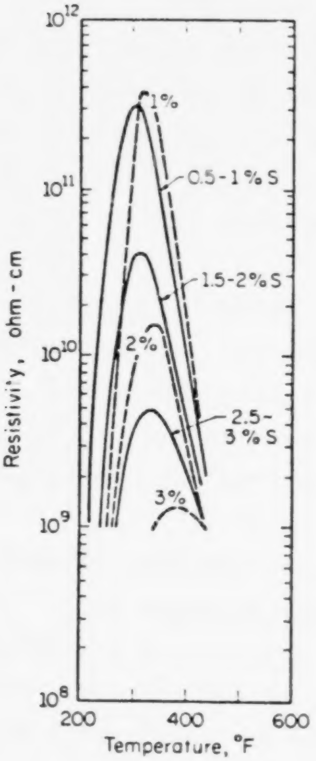

Figure 9-40. Resistivity of fly ash.*

*From Item 2, p. 188 in reference list

0.025 to 0.04 lb/ton of coal.*

- Coal crushing, screening, and conveying operations, estimated to be 0.05 lb/ton of coal.*

Dust emissions from coal storage areas are not expected to have a very strong impact on the environment because of the relatively large particle sizes encountered. Only particles having a diameter less than 30 microns have long range drift potential (greater than 1,000 feet). A very small percentage of the coal in storage is expected to be in this size range (3, p.IV-16) and therefore, dust from this source is expected to have only a localized impact, for the most part.

Control methods for particulates originating from handling and preparation activities are the following (4, p.22):

- Enclosing and ventilating screening and coal fines cleaning operations and employing wet scrubbers and bag houses to collect particulates from the ventilation air stream
- Using covered conveyors
- Reducing the stacker conveyor height to minimize the free fall of coal onto the storage pile
- Using water sprays containing a wetting agent at all transfer points, truck dump hoppers, crushers, and screens
- Employing completely enclosed coal storage facilities

The amount of fly ash in the exhaust gases from the power boiler is dependent on the boiler type, as shown by the following data (3, p.IV-9) for uncontrolled bituminous coal boilers:

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*By Wyoming Coal Gas Co. as reported in reference 4, p.22.

COAL ASH APPEARING IN BOILER EXHAUST GASES AS PARTICULATES

Boiler Type	Average as Particulates
Pulverized - Dry Bottom	85%
- Wet Bottom	65
Cyclone	20

The particle size distribution of the fly ash from the uncontrolled bituminous coal boilers is reported (3, p.IV-9) as follows:

SIZE DISTRIBUTION OF PARTICULATES FROM UNCONTROLLED BOILERS

Particle	Pulverized Type		Cyclone Type	
Size, Microns	Range*	Average*	Range*	Average*
5	12-32	25	20-72	40
10	28-56	42	40-96	65
20	42-79	65	70-99	81
40	61-93	81	81-100	92

*Weight percent less than size

Control of the fly ash in the boiler exhaust is accomplished by cyclones followed by ESP's.

Dust from ash handling is a problem only when the ash is dry, as in operation of some types of power boilers, since gasifier ash is water-quenched. Handling all ash as moist solids or as a slurry appears to be the most practical and economical means of control of particulates.

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Control of Cooling Tower Drift

Cooling tower drift is defined as mechanically entrained water droplets which are carried along with the air flowing through the tower and exhausted to the atmosphere. These water droplets have essentially the same chemical composition as the circulating water in the cooling tower.

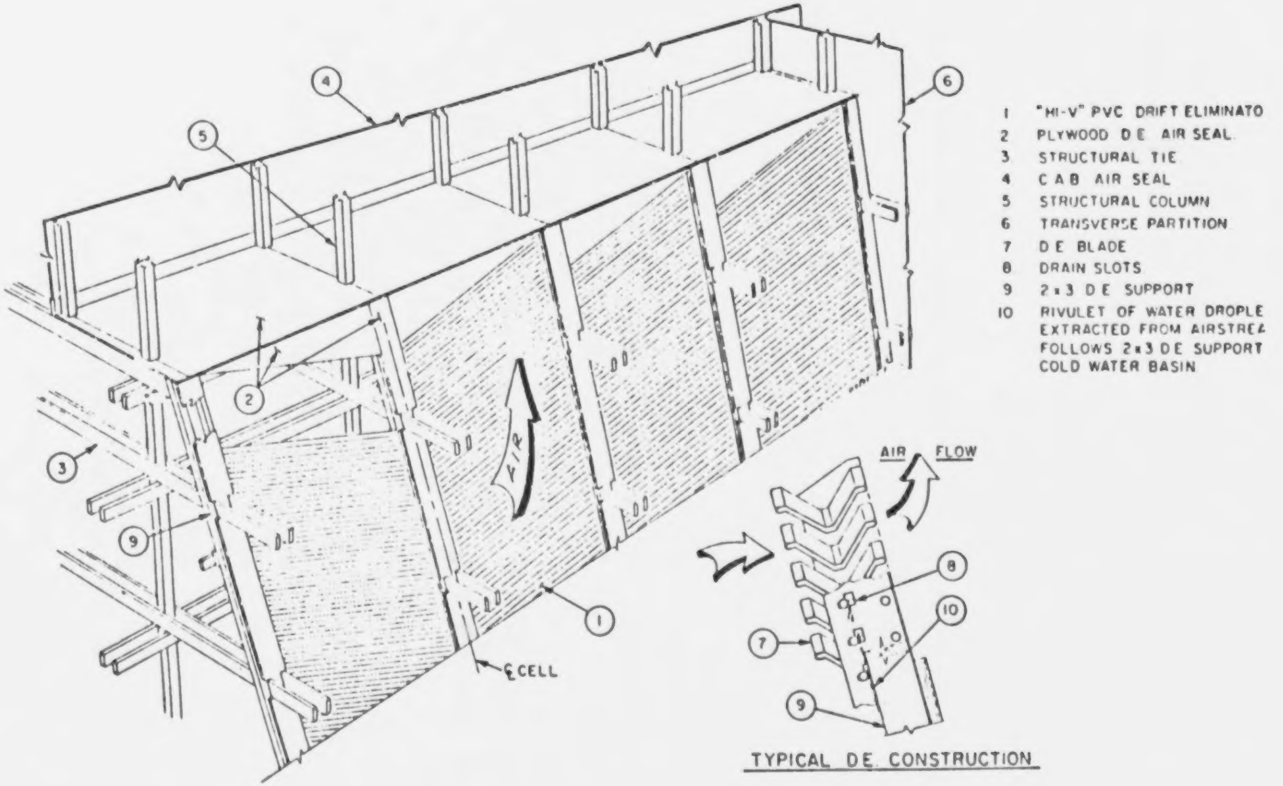
Drift resulting in deposition of water and its impurities on objects in the vicinity of the tower is potentially objectionable because corrosion problems may result, electrical equipment may be damaged, a public nuisance may be caused and vegetation may be damaged, especially with salt or brackish water cooling towers.

To determine the environmental significance of drift, it is first necessary to establish the total drift emission rate as well as the drift particle size and mass distribution. A drift measurement system, including a cyclone separator to collect entrained water droplets and an isokinetic sampling device, has been developed by Ecodyne Cooling Products Company.

To minimize the impact of drift on the environment, it is desirable to reduce its quantity. Several designs are now available to accomplish this result, of which the system developed by Ecodyne (5, pp.10,14) is chosen as an example. Ecodyne incorporates:

- A new drift eliminator design, the "Hi-V"
- Provision for positive sealing at structural members
- A positive drainage system for collected water droplets.

Figure 9-41 is a sketch of the Ecodyne Hi-V drift eliminator system.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4
5
1 "HI-V" PVC DRIFT ELIMINATO
2 PLYWOOD D E AIR SEAL
3 STRUCTURAL TIE
4 C A B AIR SEAL
5 STRUCTURAL COLUMN
6 TRANSVERSE PARTITION
7 D E BLADE
8 DRAIN SLOTS
9 2 x 3 D E SUPPORT
10 RIVULET OF WATER DROPLE EXTRACTED FROM AIRSTREA FOLLOWS 2 x 3 D E SUPPORT COLD WATER BASIN
2
AIR
AIR FLOW
3
9
8
10
1
7
℄ CELL
9
TYPICAL D E CONSTRUCTION

Figure 9-41. Sketch of Ecodyne "Hi-V" drift eliminator system*

*From Item 5, p.10 in reference list

Drift tests have been conducted recently by Ecodyne on industrial cooling towers. Both standard towers, equipped with the two-pass drift eliminator configurations that are typical of the industry for the past twenty years, and towers equipped with the Ecodyne Hi-V system were tested. The results of 22 drift tests are as follows:

	Standard	Hi-V
Drift, % of Circulation		
Range	0.02-0.12	0.001-0.008
Typical Value	0.05	0.004
Particle Size Range, microns	22-2,400	22-2,400

These tests indicate that drift can be reduced by over 90 percent through use of the Ecodyne system, or one having similar effects.

The dynamic behavior of drift is a function of the original droplet size, condensation or evaporation rates, aerodynamic and gravitational forces, and meteorological conditions. Ecodyne developed a computer program to evaluate the trajectory path of drift droplets. The program takes into account the change in drop size and position with respect to time on an incremental basis, together with such factors as original drop size, tower operating parameters, atmospheric conditions, fall velocity and evaporation rates. TABLE 9-11 shows the size and mass distribution of drift particles. Figure 9-42 shows the fall velocity of water drops as a function of size.

An eight cell crossflow tower designed to cool 134,000 GPM of salt water with the same chemical composition and salinity level as the sea was chosen for an example. The plant location is assumed to be two miles from the ocean on an estuary or bay. The drift rate is 0.004 percent of the circulating water rate and

TABLE 9-11. SIZE AND MASS DISTRIBUTION OF DRIFT PARTICLES (HI-V DRIFT ELIMINATORS)

DROPLET DIAMETER (MICRON)	% OF SAMPLE BY NUMBER	% MASS BY DROPLET SIZE
22	24.0	0.43
29	36.0	1.49
44	26.0	3.76
58	6.3	2.09
65	4.0	1.86
87	1.4	1.56
108	0.67	1.43
120	0.43	1.26
132	0.28	1.09
144	0.26	1.32
174	0.65	5.81
300	0.11	5.04
450	0.027	4.17
600	0.011	4.01
750	0.0055	4.00
900	0.0033	4.03
1050	0.0024	4.57
1200	0.0019	5.46
1350	0.0016	6.80
2250	0.00095	17.99
2400	0.0010	21.83

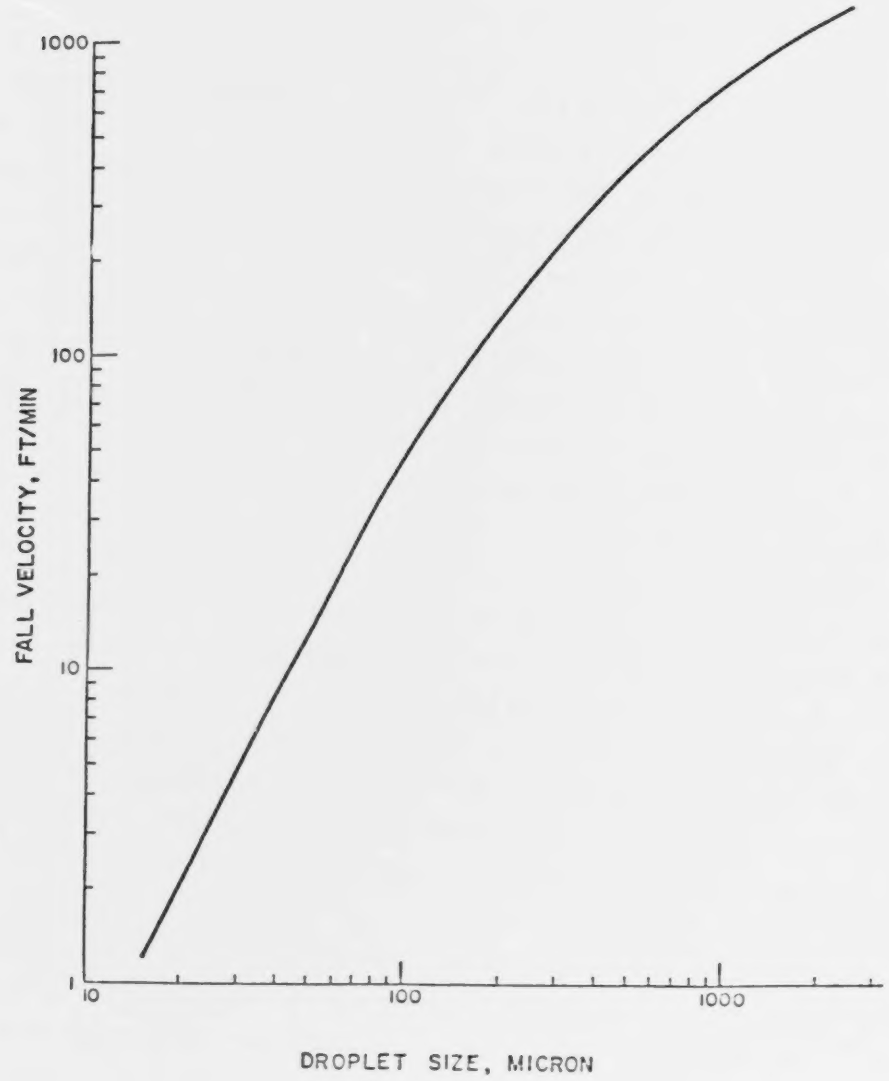


Figure 9-42. Fall velocity of water drops as a function of size.*

*From Item 5 in reference list

the drift mass size distribution of TABLE 9-11 is used in this example. The conditions were established as 17°C dry bulb temperature and 50 percent relative humidity with stable atmospheric conditions and an onshore wind of 20 miles per hour at the plant location.

Figure 9-43 illustrates the drift dynamic behavior for this example. The results show that 68.7 percent of the drift mass hits the ground in the first 400 feet. Note that roughly half of this total mass falls out in the first 150 feet. Drift originally smaller than 100 microns in diameter represents 12 percent of the total mass and, for the atmospheric conditions chosen, droplets originally smaller than 450 microns evaporate to a diameter of 100 microns or less before hitting the ground. Thus 31.3 percent of the total drift mass remains in suspension and its subsequent behavior is evaluated using an atmospheric dispersion model.

The dashed lines of Figure 9-43 show the trajectory history when the atmospheric conditions are changed to -1°C dry bulb temperature and 80 percent relative humidity. Note that for this typical winter condition the droplets fall slightly closer to the tower since the evaporation rate is diminished. Further, there is a slight change in the total airborne drift quantity since the 450 micron droplets now hit the ground.

A parametric analysis, where relative humidity, salinity levels, and atmospheric and exit air temperatures were varied, revealed that the dynamic behavior of drift droplets in the 400 to 700 micron size range is the most sensitive to changes in these variables. This example shows, however, that those drift droplets that hit the ground will do so in the first 500 feet with a 20 mph wind.

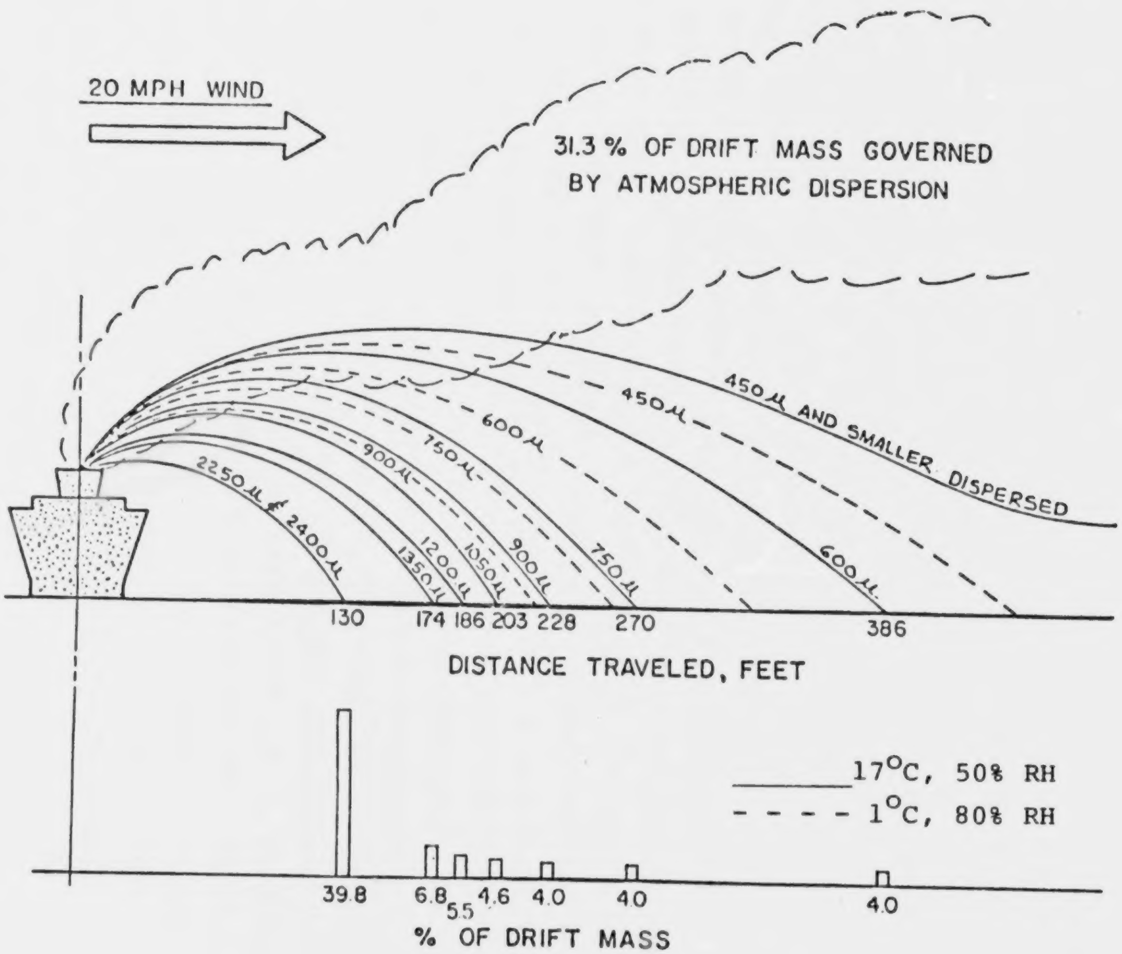


Figure 9-43. Dynamic behavior of cooling tower drift.*

*From Item 5 in reference list

Evaluation of other conditions including weather extremes show that even under the most adverse conditions all drift droplets that will reach the ground will do so in the first 1,000 feet.

The relative significance of the airborne drift in terms of salt concentrations must now be determined. Natural airborne salt nuclei are generated by the bursting of air bubbles on the surface of the sea. This is caused by wind, waves, and surf action. Meteorologists have been actively interested in these sea salt nuclei and their role in droplet formation in clouds and precipitation for many years. It has been shown that wind speed, direction, and the distance from the shore line will determine the natural sea salt nuclei concentration levels. Typical values are shown in Figure 9-44. The air salt concentration is shown to be directly proportional to inblowing wind speed and inversely proportional to the distance from the shoreline. Note that the airborne sea salt concentrations two miles from the shoreline vary anywhere from 9_{μ} g/m^{3} to over 150μg/m^{3} for normal wind speed variations. Obviously local plant life in the area must be capable of withstanding these natural airborne salt levels if they are to survive. A recent comprehensive environmental report (7) established an important correlation between airborne salt concentration levels and injury to vegetation. Based on field observations it was shown that exposure of local vegetation to airborne salt concentrations above 100_{μ} g/m^{3} for several hours would result in some foliar injury. There was no visible damage for concentration levels below 60_{μ} g/m^{3}. This information suggests that a conservative plant damage threshold level can be established at 60_{μ} g/m^{3}. For the example chosen, the background natural airborne sea salt concentration is 47_{μ} g/m^{3} (Figure 9-44) and application of a dispersion model shows that the total of the airborne drift plus background concentration will reach 60_{μ} g/m^{3} at a position approximately 2,200 feet downwind from the tower. Therefore no plant damage is anticipated beyond this distance.

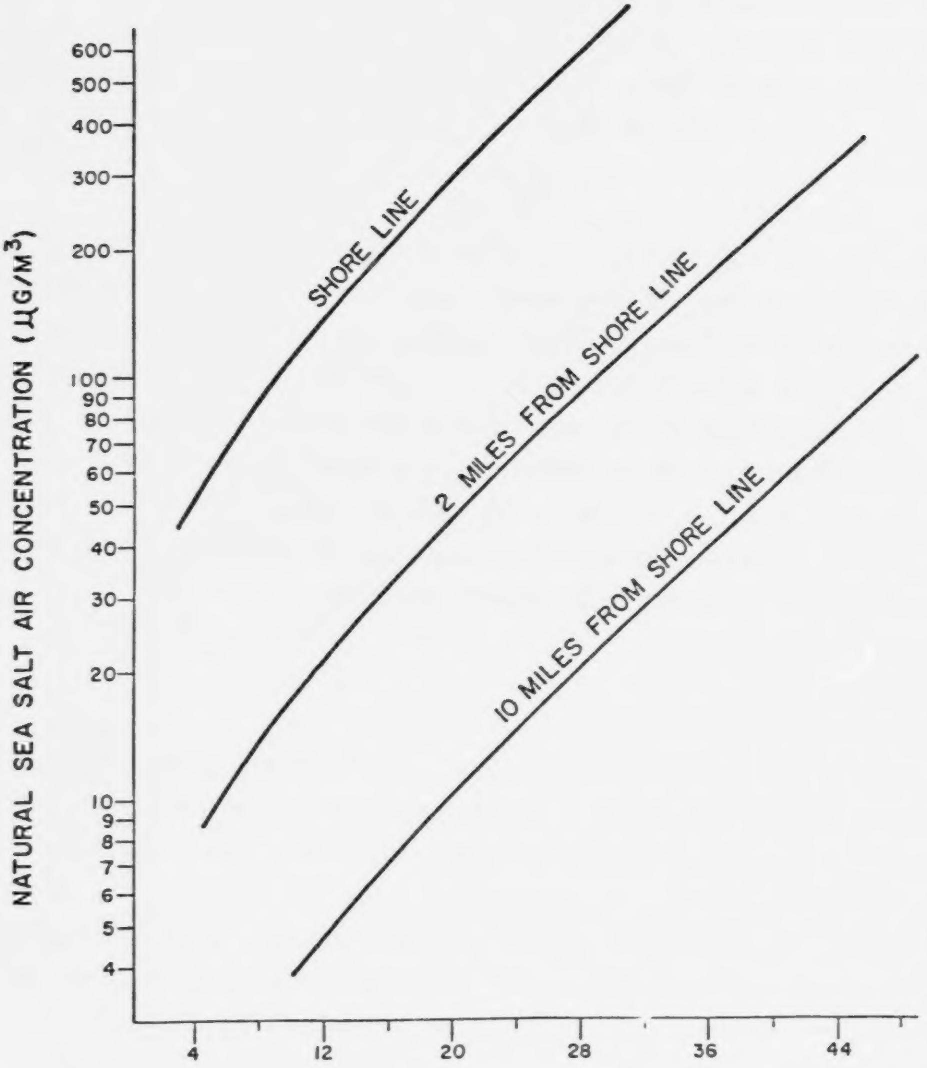


Figure 9-44. Natural sea salt concentration in air.*

*From Item 5 in reference list

Note also that the drift-related increase is 13_{μ} g/m^3 at this position. Figure 9-44 shows that a 13_{μ} g/m^3 change in airborne salt level is approximately equivalent to a 3 mph change in average wind speed. Thus the fractional increase in airborne salt concentrations due to drift is insignificant when compared to normal variations caused by changes in atmospheric wind conditions.

In summary, it can be generally concluded that cooling tower drift effects on the environment are localized and that beyond some reasonable distance, that is usually within the plant site boundary, drift does not significantly affect the environment. All field experience during the last 20 years where salt or brackish water has been used in cooling towers supports this general conclusion. Further, it can be concluded that the use of modern drift eliminator systems can reduce drift losses by about 90 percent when compared to older systems.

References--

1. Wistrom, G.K., and Ovard, J.C., "Cooling Tower Drift, Its Measurement, Control, and Environmental Effects." Cooling Tower Institute Annual Meeting, Houston, Jan. 1973. 893*

2. Chilton, H., "Elimination of Carryover from Packed Towers with Special Reference to Natural Draught Water Cooling Towers." Trans. Institution of Chemical Engineers, Vol. 30, 1952.

3. "Assessment of Environmental Effects." General Public Utilities Unpublished Report, January 1972.

*Pullman Kellogg Reference File Number

Miscellaneous Control Techniques

Control of Lock Hopper Vent Gases--

In the Lurgi process, as well as in certain others, coal is fed to the gasifiers in a cyclic operation using a pressurized lock hopper. The recommended sequence of operations is as follows:

- Fill the lock hopper to about 90 percent capacity with coal from coal bunker
- Pressurize the lock hopper with cooled crude gas
- Feed coal from the lock hopper to the gasifier
- As the pressure tends to drop in the lock hopper, add cooled crude gas by use of a recycle compressor or, as an alternate, gas from the top of the gasifier can back flow through the entering coal stream for pressure equalization
- When the lock hopper has been emptied of coal, bleed the gas into the fuel system or return it to the process by means of a compressor
- Displace the residual gas in the lock hopper with CO_2 or nitrogen and direct this stream to the fuel system or to incineration
- Open the lock hopper and recharge it with coal

The use of the scheme recommended prevents the discharge of pollutants (CO, H_2S, COS, hydrocarbons) to the atmosphere while recovering the fuel value of the lock hopper pressurizing gas.

Control of Ash Quench Vent--

Ash produced in the Lurgi gasifier is discharged through the bottom of the gasifier into a pressurized ash lock. After the ash lock is filled with ash, the top ash lock cone valve is closed, isolating the ash lock chamber. High pressure gases in the ash lock at this point are mainly steam. The chamber is vented to a close coupled direct contact condenser, where the steam is condensed with a water spray. The bottom ash lock valve is then opened and the ash falls out. After the ash is dumped, both cone valves are closed and the ash lock chamber is repressurized with steam. The top ash lock valve is opened and ash flow from the producer is re-established. The lock dumps approximately every 20 minutes into an ash quench system where a mixture of water streams from the plant are added. During the quenching process a large amount of steam containing fine ash dust and clinkers is produced.

The recommended control method for disposal of this stream is to send the mixture first to a wet cyclone to remove the clinkers and then to a direct contact condenser to condense the steam and remove fine ash particles. Along with the steam, some amount of non-condensable gases (hydrocarbons) may be formed from organic materials in the quench water and unreacted carbon in the ash. The quantity and composition of this gas stream is not known precisely; however, data from SASOL reveals that the gas is mostly steam. Nevertheless, final treatment by incineration is recommended.

Control of Miscellaneous Leaks--
In any processing plant, certain leaks will arise from valve stems, pump packing glands and mechanical seals, flanges, relief valves, instrument and piping connections,and compressor seals. In most cases, emissions from these sources can be reduced or eliminated by proper plant design and thorough maintenance programs.

Some specific methods of control are the following:

- Provide enclosures for valve stems, pump seals, compressor seals, flanges, and instrument and piping connections. Escaping gases can be collected in a central system, or systems, and routed to the incinerator

- Provide a closed relief valve system which vents to the incinerator

- Provide a positive pressure sealing system for compressors in which an inert gas, such as nitrogen, can provide positive pressure in the outer seal chamber. Process gas which passes the inner seal mixes with the inert gas and is bled to the incinerator through an intermediate seal chamber
- Provide covers and collection systems for escaping hydrocarbon vapors from open systems such as API separators

Control of Emissions from Storage Vessels--
In a coal conversion facility a number of products, byproducts and chemicals must be stored at the site. Some of these are tar, tar oils, phenols, ammonia, naphtha, sulfur, sodium hydroxide, sulfuric acid, methanol, isopropyl ether, Stretford solution components, Selexol solvent, and water treating chemicals.

Emissions from storage will consist of tank breathing, leaks, spills, venting of tanks while filling and vaporization (boil-off) of volatile materials.

Liquid spills are handled best by providing containing dikes around the storage vessels to prevent spreading of spilled liquid while recovery or disposal operations are carried out.

The means of control for emissions depends on the vapor source:
- Vapor collection systems for atmospheric storage tanks should be so designed that then filling these tanks the displaced vapors will be diverted to the fuel system, incinerator or recovery systems. Secondary wiper seals to reduce hydrocarbon losses from floating roof tanks should be provided
- Refrigeration systems will control boil-off from cryogenic

or volatile liquid storage

- Scrubbing systems which use low volatility solvents to absorb escaping vapors may be provided
- Adsorption systems using activated carbon or other suitable materials may be provided as first stage or backup methods to prevent the escape of storage vapors.

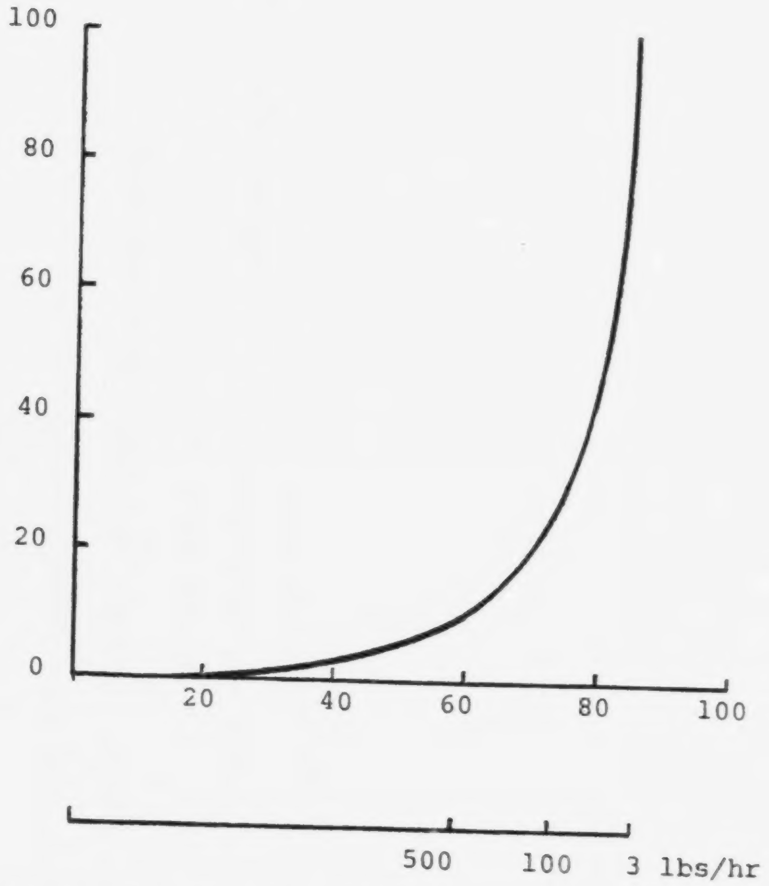
Cost of Emission Control--

It is difficult, if not impossible, to quantify the cost of emission control for the miscellaneous sources previously mentioned. If a choice existed, control measures would be applied first to those sources which contributed the most emissions. It is interesting to note that a DuPont study (1) concluded that:

- Almost 75 percent of inventoried organic emissions came from sources which emit more that 500 lb/hr and these could be reduced by 85 percent for about 10 percent of the investment cost that would be needed for controlling all sources which emit more than 3 lb/hr. See Figure 9-45.
- About 95 percent of emissions come from sources which emit more than 23 lb/hr. Figure 9-45 indicates that these could be reduced by 85 percent for about 50 percent of the investment cost that would be needed for controlling all sources which emit more than 3 lb/hr.
- Figure 9-46 demonstrates that the operating cost of emission control on a unit basis begins to increase rapidly for those sources which emit less than 100 lb/hr. It increases very rapidly for sources emitting less than 20 to 30 lb/hr.

One may conclude from the above information that:

- First priority for control should be given to the sources with the greatest emission rates. This conclusion is reached from consideration of both environmental impact and economics.



BASIS: 100% INVESTMENT - $350MM (1976 DOLLARS)
Sources > 3 lb/hr or 15 lb/day

ALL SOURCES GREATER THAN INDICATED EMISSION RATE CONTROLLED TO 85%

Figure 9-45. Investment vs. emission reduction.*

*From Item 1 in reference list

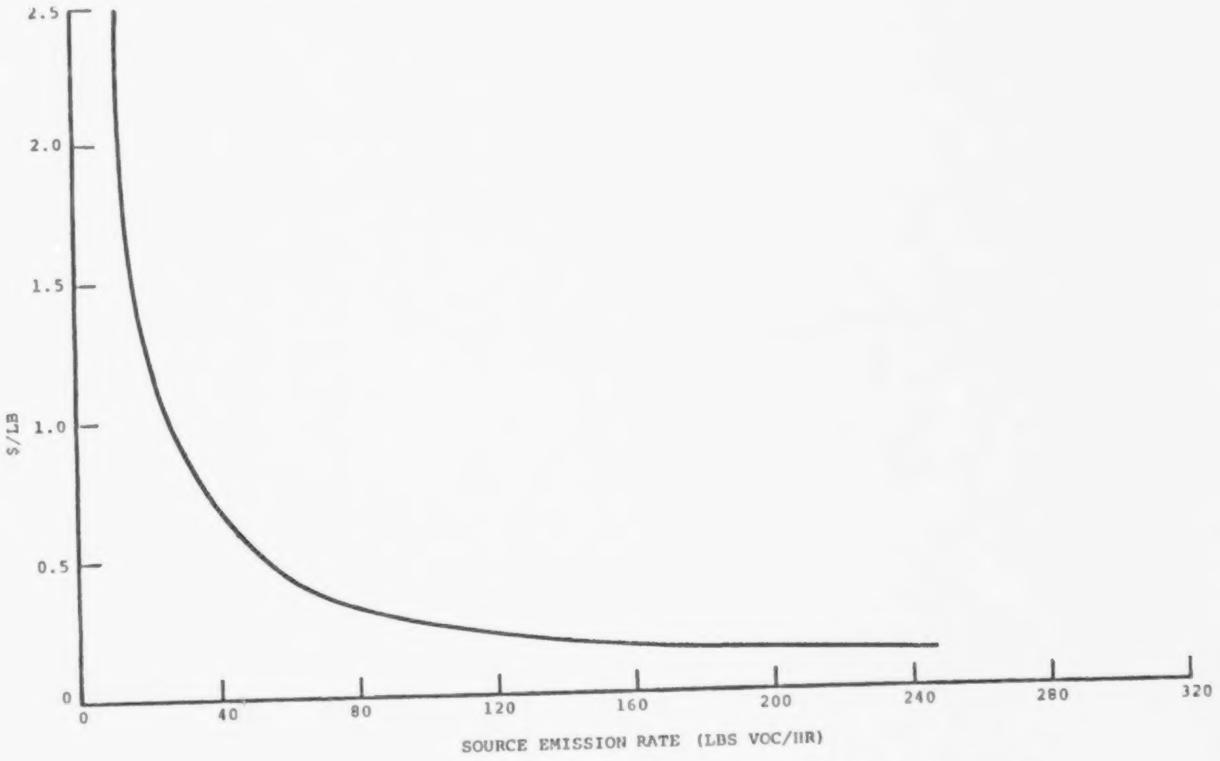


Figure 9-46. Organic abatement operating cost.*

*From Item 1 in reference list.

- Means to reduce the unit cost of control for small emissions should be developed, such as central collection systems which tie in to a number of emission sources.
- Some consideration might be given to relaxing regulations as far as small emissions are concerned. An emission source of 200 lb/hr which is reduced by 85 percent results in a final emission of 30 lb/hr. This is equal to 10 small sources, each emitting 3 lb/hr. Requiring emission control systems for all 10 small sources may not be entirely practical because of the limited effectiveness and high cost of these systems.

Reference--

1. Kittleman, T., and Akell, R., "Cost of Alternative Organic Emission Control Regulations." AIChE Meeting, New York, Nov. 1977. 860*

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*Pullman Kellogg Reference File number

INTEGRATED SCHEMES FOR EMISSIONS CONTROL

The Lurgi Dry Ash process was selected as the base gasification case for study of integrated schemes for emissions control. As previously mentioned, the block flow diagram and material balance of Figure 9-1 was assembled from the conceptual designs of C. F. Braun, Cameron Engineers and Pullman Kellogg for operation of the Lurgi process on western, low sulfur coal. From this flow diagram and material balance the base case flow diagrams for the sections of the plant were developed as shown in Figures 9-2 through 9-9, then the overall feed and product weight balance of Figure 9-10 and the gaseous emission streams shown in Figure 9-11 were calculated. These flow diagrams, calculations and material balances were primarily directed toward establishment of the operating characteristics of the sulfur recovery unit, the composition of the offgas stream from sulfur recovery and the required operating characteristics of the unit for control of the sulfur emissions from the coal conversion plant. The overall sulfur balance for the Lurgi base case is shown in Figure 9-47. The changes in the flows and the material balance when the Lurgi process is fed with high sulfur coal are shown in Figure 9-48.

The Lurgi Dry Ash process produces phenols, oils and tars (p/o/t) that are separated from the gas stream and either processed further for sale or are sent to an incinerator/boiler. Because of the quantity of materials that must be disposed of by incineration, the incinerator/boiler is an important part of the coal conversion plant, its operation is closely integrated with the sulfur recovery unit and treatment of its offgases provides a second source of product sulfur.

On the other hand, the high temperature gasification processes, exemplified by the Bi-Gas process, produce little or no p/o/t and the importance of the incinerator/boiler as a means of waste

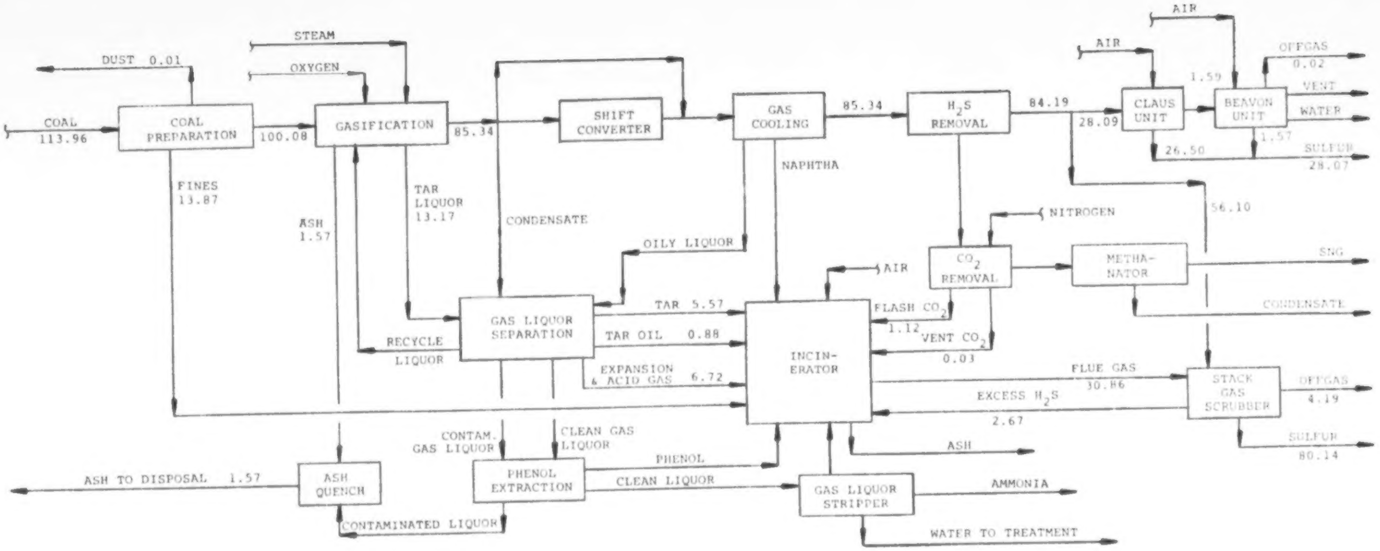
COAL 113.96
DUST 0.01
COAL PREPARATION
100.08
FINES 13.87
STEAM
OXYGEN
GASIFICATION
85.34
ASH 1.57
TAR LIQUOR 13.17
RECYCLE LIQUOR
SHIFT CONVERTER
CONDENSATE
GAS COOLING
85.34
NAPHTHA
OILY LIQUOR
H_2S REMOVAL
84.19
28.09
AIR
CLAUS UNIT
26.50
AIR
1.59
BEAVON UNIT
1.57
OFFGAS 0.02
VENT
WATER
SULFUR 28.07
56.10
NITROGEN
CO_2 REMOVAL
METHA-NATOR
SNG
CONDENSATE
AIR
GAS LIQUOR SEPARATION
TAR 5.57
TAR OIL 0.88
EXPANSION & ACID GAS 6.72
FLASH CO_2 1.12
VENT CO_2 0.03
INCIN-ERATOR
FLUE GAS 30.86
EXCESS H_2S 2.67
STACK GAS SCRUBBER
OFFGAS 4.19
SULFUR 80.14
ASH
CONTAM. GAS LIQUOR
CLEAN GAS LIQUOR
PHENOL EXTRACTION
PHENOL
CLEAN LIQUOR
GAS LIQUOR STRIPPER
AMMONIA
WATER TO TREATMENT
ASH TO DISPOSAL 1.57
ASH QUENCH
CONTAMINATED LIQUOR

Figure 9-47.- Sulfur balance : Lurgi gasification base case with low sulfur coal. (Sulfur flows in tons per day.)

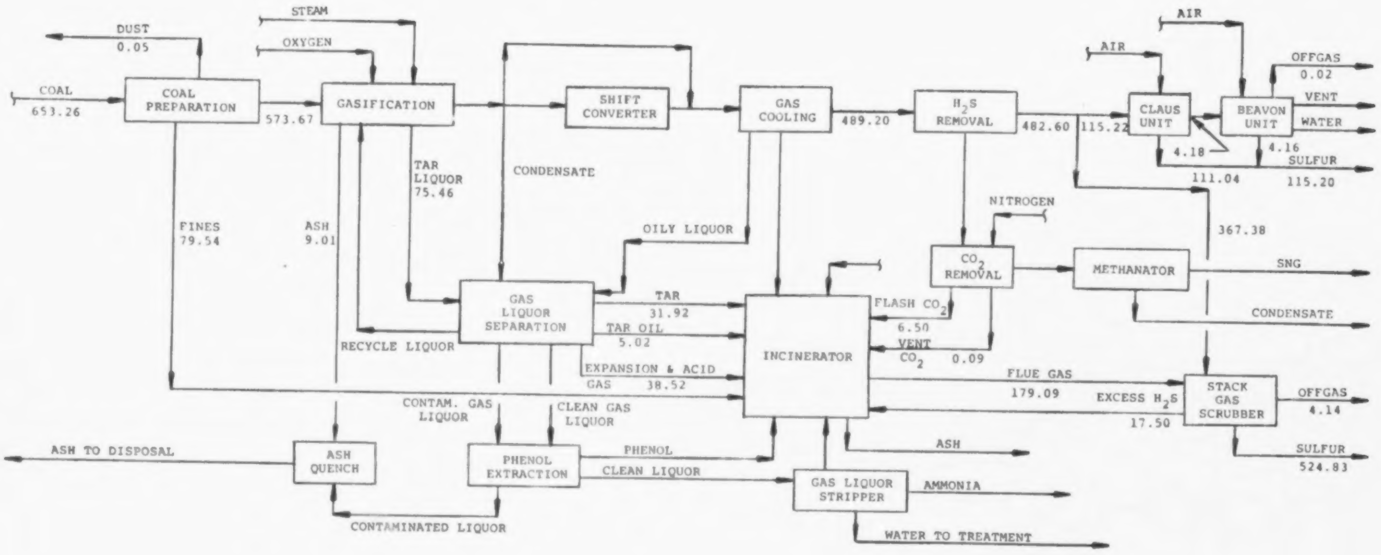
COAL
653.26
DUST
0.05
COAL PREPARATION
573.67
STEAM
OXYGEN
GASIFICATION
SHIFT CONVERTER
GAS COOLING
489.20
H2S REMOVAL
482.60
115.22
AIR
AIR
CLAUS UNIT
BEAVON UNIT
OFFGAS
0.02
VENT
WATER
4.18
4.16
SULFUR
111.04
115.20
TAR LIQUOR
75.46
CONDENSATE
FINES
79.54
ASH
9.01
OILY LIQUOR
NITROGEN
367.38
CO2 REMOVAL
METHANATOR
SNG
CONDENSATE
GAS LIQUOR SEPARATION
TAR
31.92
TAR OIL
5.02
EXPANSION & ACID GAS
38.52
RECYCLE LIQUOR
INCINERATOR
FLASH CO2
6.50
VENT CO2
0.09
FLUE GAS
179.09
EXCESS H2S
17.50
STACK GAS SCRUBBER
OFFGAS
4.14
SULFUR
524.83
CONTAM. GAS LIQUOR
CLEAN GAS LIQUOR
ASH TO DISPOSAL
ASH QUENCH
PHENOL EXTRACTION
PHENOL
CLEAN LIQUOR
GAS LIQUOR STRIPPER
ASH
AMMONIA
WATER TO TREATMENT
CONTAMINATED LIQUOR

Figure 9-48.- Sulfur balance : Lurgi gasification with high sulfur coal.
(Sulfur flows in tons per day.)

disposal is reduced. Process steam is raised more from coal and less from waste, the overall sulfur balance changes and the demands on the sulfur recovery and emissions control units change. The overall sulfur balance for Bi-Gas operation was based on the C. F. Braun conceptual design for operation with low sulfur coal and is shown in Figure 9-49. The changes in the sulfur balance when high sulfur coal is fed to the process are illustrated in Figure 9-50.

For liquefaction, the Ralph M. Parsons conceptual design for the SRC II process was selected as representative. As previously discussed, the liquefaction flow diagram and material balance are shown in Figure 9-12 and those for sections of the plant are shown in Figures 9-13 through 9-20. From these diagrams the overall feed and product weight balance shown in Figure 9-21 was calculated and the gaseous emission streams shown in Figure 9-22 were established. As in the gasification processes, calculation of the material balances was primarily directed toward determining the operating characteristics of the sulfur recovery unit, the offgas stream composition and the demands on the unit for control of the sulfur emissions. The overall sulfur balance for liquefaction is shown in Figure 9-51.

The properties of the low and high sulfur coals that are considered as the feeds to the illustrated processes are shown in TABLE 9-12.

Incineration and Steam Generation

Processes Producing Phenols, Oils and Tars: The Base Case--

In the Lurgi process, chosen as being typical of those conversion processes producing phenols, oils and tars (p/o/t) in the gasifier, steam is required for the process and for generation of the electric power (about 350,000 horsepower for production of

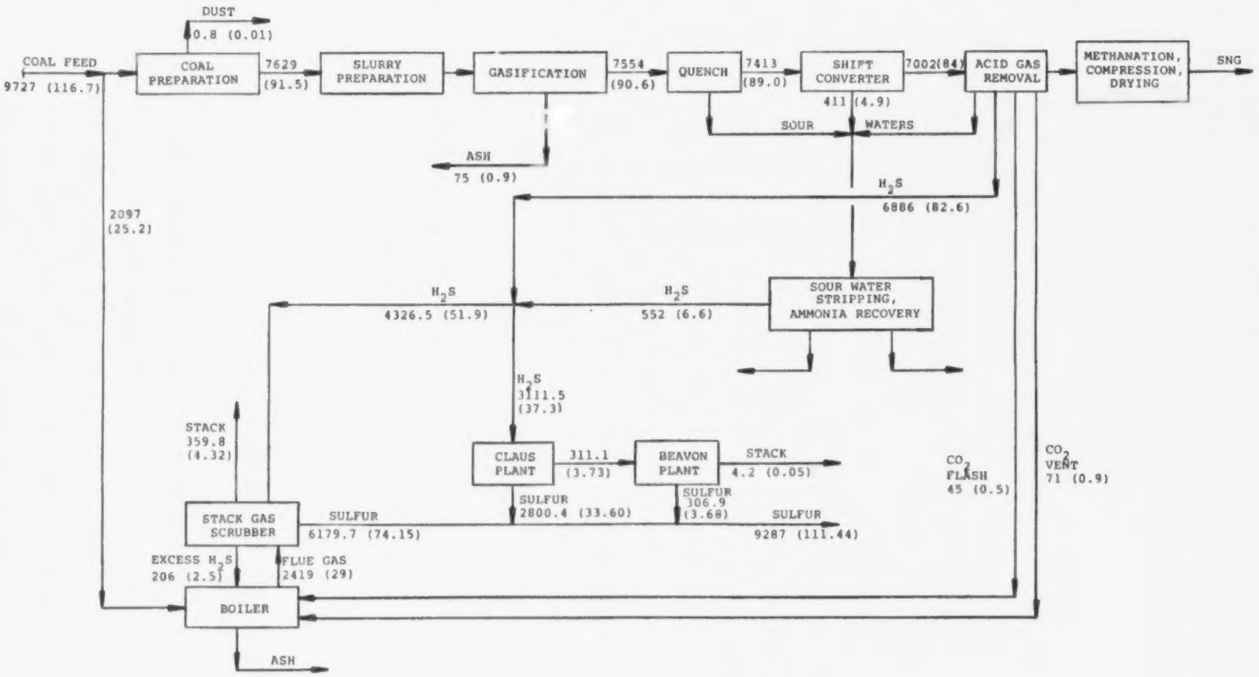
COAL FEED
9727 (116.7)
DUST
0.8 (0.01)
COAL PREPARATION
7629 (91.5)
SLURRY PREPARATION
GASIFICATION
7554 (90.6)
QUENCH
7413 (89.0)
SHIFT CONVERTER
7002(84)
ACID GAS REMOVAL
METHANATION, COMPRESSION, DRYING
SNG
ASH
75 (0.9)
411 (4.9)
SOUR
WATERS
H2S
6886 (82.6)
2097 (25.2)
H2S
4326.5 (51.9)
H2S
552 (6.6)
SOUR WATER STRIPPING, AMMONIA RECOVERY
H2S
3111.5 (37.3)
CLAUS PLANT
311.1 (3.73)
BEAVON PLANT
STACK
4.2 (0.05)
SULFUR
2800.4 (33.60)
SULFUR
306.9 (3.68)
SULFUR
9287 (111.44)
CO2 FLASH
45 (0.5)
CO2 VENT
71 (0.9)
STACK
359.8 (4.32)
STACK GAS SCRUBBER
SULFUR
6179.7 (74.15)
EXCESS H2S
206 (2.5)
FLUE GAS
2419 (29)
BOILER
ASH

Figure 9-49.- Sulfur balance: Bi-Gas process with low sulfur coal.

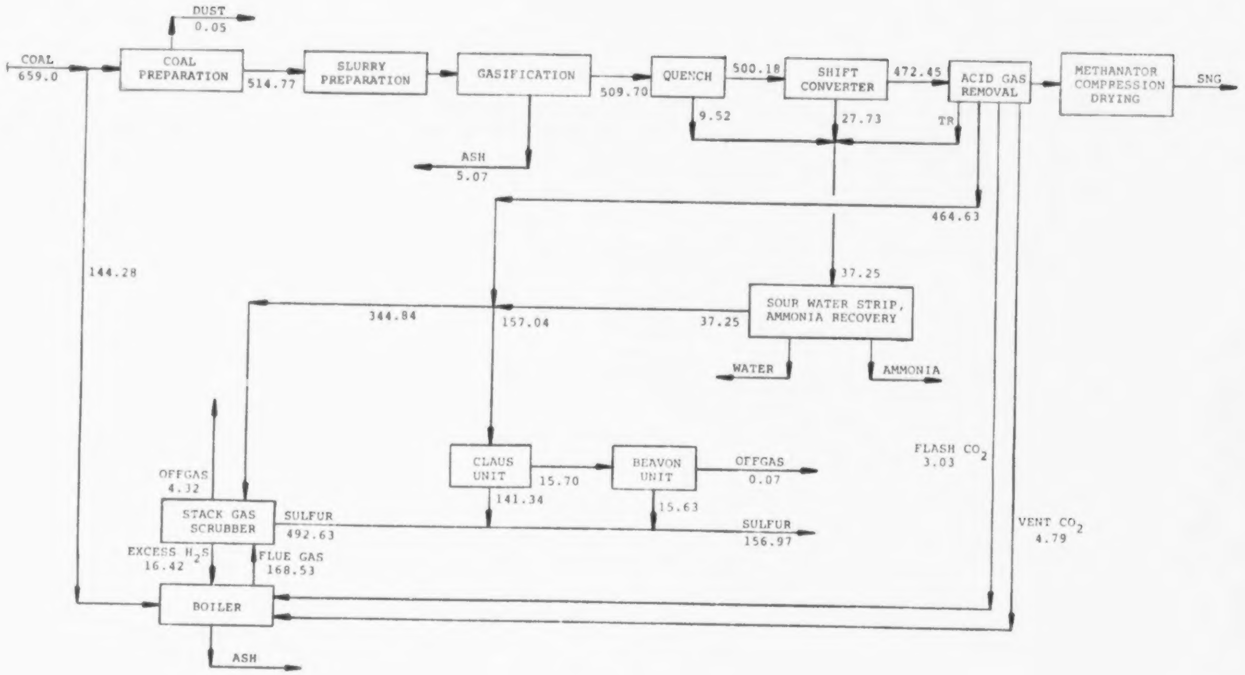
COAL
659.0
DUST
0.05
COAL PREPARATION
514.77
SLURRY PREPARATION
GASIFICATION
509.70
QUENCH
500.18
SHIFT CONVERTER
472.45
ACID GAS REMOVAL
METHANATOR COMPRESSION DRYING
SNG
9.52
27.73
TR
ASH
5.07
464.63
144.28
37.25
SOUR WATER STRIP, AMMONIA RECOVERY
344.84
157.04
37.25
WATER
AMMONIA
FLASH CO2
3.03
CLAUS UNIT
15.70
BEAVON UNIT
OFFGAS
0.07
141.34
15.63
OFFGAS
4.32
STACK GAS SCRUBBER
SULFUR
492.63
SULFUR
156.97
VENT CO2
4.79
EXCESS H2S
16.42
FLUE GAS
168.53
BOILER
ASH

Figure 9-50.- Sulfur balance: Bi-Gas gasification with high sulfur coal.

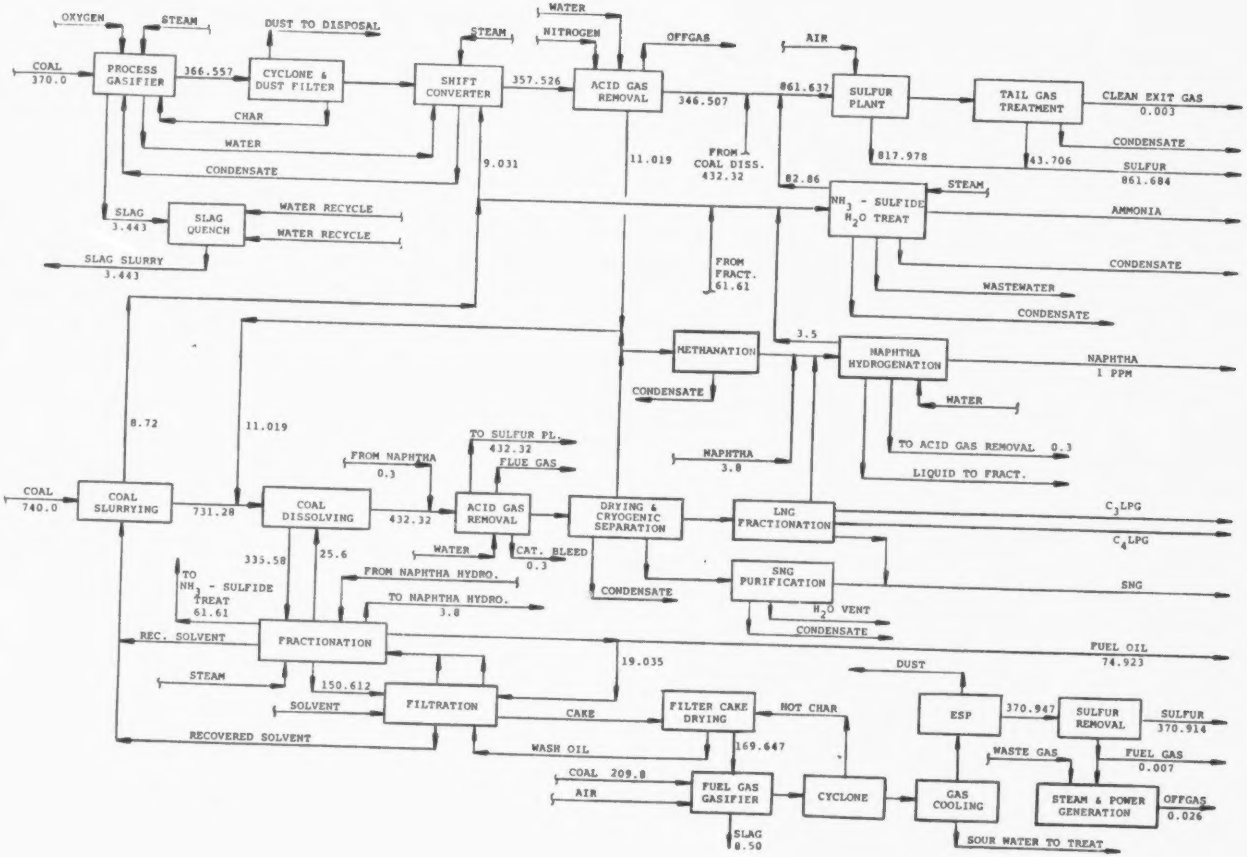
OXYGEN
STEAM
COAL
370.0
PROCESS GASIFIER
366.557
CYCLONE & DUST FILTER
DUST TO DISPOSAL
CHAR
WATER
CONDENSATE
STEAM
SHIFT CONVERTER
357.526
9.031
WATER
NITROGEN
ACID GAS REMOVAL
OFFGAS
346.507
11.019
FROM COAL DISS. 432.32
861.637
AIR
SULFUR PLANT
817.978
82.86
TAIL GAS TREATMENT
CLEAN EXIT GAS 0.003
CONDENSATE
43.706
SULFUR 861.684
SLAG 3.443
SLAG QUENCH
WATER RECYCLE
WATER RECYCLE
SLAG SLURRY 3.443
NH_3 - SULFIDE H_2O TREAT
STEAM
AMMONIA
CONDENSATE
WASTEWATER
CONDENSATE
FROM FRACT. 61.61
METHANATION
CONDENSATE
3.5
NAPHTHA HYDROGENATION
NAPHTHA 1 PPM
WATER
TO ACID GAS REMOVAL 0.3
LIQUID TO FRACT.
NAPHTHA 3.8
8.72
11.019
COAL 740.0
COAL SLURRYING
731.28
COAL DISSOLVING
FROM NAPHTHA 0.3
432.32
TO SULFUR PL. 432.32
FLUE GAS
ACID GAS REMOVAL
WATER
CAT. BLEED 0.3
DRYING & CRYOGENIC SEPARATION
CONDENSATE
LNG FRACTIONATION
C_3LPG
C_4LPG
SNG PURIFICATION
SNG
H_2O VENT
CONDENSATE
335.58
25.6
TO NH_3 - SULFIDE TREAT 61.61
FROM NAPHTHA HYDRO.
TO NAPHTHA HYDRO. 3.8
REC. SOLVENT
FRACTIONATION
FUEL OIL 74.923
19.035
STEAM
150.612
SOLVENT
FILTRATION
CAKE
RECOVERED SOLVENT
WASH OIL
FILTER CAKE DRYING
HOT CHAR
169.647
COAL 209.8
AIR
FUEL GAS GASIFIER
SLAG 8.50
CYCLONE
DUST
ESP
370.947
SULFUR REMOVAL
SULFUR 370.914
WASTE GAS
FUEL GAS 0.007
GAS COOLING
STEAM & POWER GENERATION
OFFGAS 0.026
SOUR WATER TO TREAT

Figure 9-51.- Sulfur balance: SRC-II liquefaction.

TABLE 9-12. COAL PROPERTIES

	To Gasification		
	Western Coal	Eastern Coal	To Liquefaction, Midwest Coal
Proximate Analysis, As Received, Wt%			
Moisture	22.0	6.0	2.7
Volatile Matter	29.4	31.9	39.7
Fixed Carbon	42.6	51.5	45.8
Ash	6.0	10.6	11.8
	100.0	100.0	100.0
Ultimate Analysis (dry), Wt%			
Carbon	67.70	71.50	69.06
Hydrogen	4.61	5.02	4.73
Nitrogen	0.85	1.23	1.34
Oxygen	18.46	6.53	8.94
Sulfur	0.66	4.42	3.80
Ash	7.72	11.30	12.13
	100.00	100.00	100.00
Heating Value of Dry Coal Btu/lb (HHV)	11,290	13,190	12,460
Heating Value of Coal as Received, Btu/lb (HHV)	8,800	12,400	12,125

250 billion Btu/day of SNG) required for the conversion plant complex. The steam generator is considered as a process auxiliary in most process designs and is usually included in the offsite section of the plant complex. In this illustrative Lurgi case, however, with operation on western coal, the heating value of the liquid byproduct and waste streams, and of the waste gas streams, amounts to about 58 percent of the total energy required for steam and power generation. Therefore, the steam generator may be considered as an incinerator and as a part of the conversion process within battery limits. The remaining 42 percent of the energy requirement is supplied by part of the total plant coal feed in the Lurgi design, but alternately may be supplied by other fuels.

In addition to providing a useful and economical means of disposal for combustible liquid and gaseous byproducts and wastes, the incinerator/boiler system aids in reducing or eliminating emissions of:

- Particulates from coal firing
- CO, hydrocarbons (CH_4, C_2H_4, C_2H_6) and ammonia from waste gas streams
- Sulfur from gas, liquid and solid (coal) feeds. Sulfur and sulfur compounds are oxidized to SO_2 (and SO_3) and then are removed in the flue gas desulfurization (FGD) unit. The FGD unit will be discussed later
- Nitrogen oxides (NO_x)

To illustrate the integration of the incinerator into the process scheme, the design conditions of the incinerator unit were taken as follows:

Heat Input	4678.5 MM Btu/Hr (HHV)
Steam Flow	2,844,000 Lb/Hr
Steam Pressure	1,500 psig at superheater outlet

Steam Temperature	910°F psig at superheater outlet
F. W. Temperature	456°F
Stack Temperature	300-350°F
Air Temperature	95°F

The design plan contemplates burning a combination of fuels in the boiler(s). Miscellaneous gas streams are to be incinerated to destroy H_2, CO, hydrocarbons,and ammonia and to oxidize sulfur compounds. Liquid byproducts are to be burned, together with coal fines to supply the remaining heat.

The gaseous fuels are described in TABLE 9-13. The liquid fuels are described in TABLE 9-14. Additional fuel required is comprised of coal fines with the ultimate analysis (dry basis) shown in TABLE 9-12 and containing 22 percent moisture.

The net coal required by the boilers is determined by deduction to be about 175,190 lb/hr (dry) or 224,603 lb/hr (wet) which furnished 1,977.9 MM Btu/hr (HHV).

The total input to the boilers is summarized as follows:

	Gas	Liquid	Coal	Total
Lb/hr	1,640,947	164,282	224,603	2,029,832
MM Btu/hr(HHV)	144.6	2,556.0	1,977.9	4,678.5
Heat input, %	3.091	54.633	42.276	100.000

Combustion air required at 15 percent excess is 160,160 moles per hour dry plus 3,994 moles of water. Total weight of air is 4,711,627 lbs/hr. The calculated flue gas composition is shown in TABLE 9-15.

Information received from Combustion Engineering indicates that three boilers are needed to satisfy the demand. Each of the units

TABLE 9-13. GASEOUS FUELS TO INCINERATOR

	Exp. Gas MPH	Acid Gas MPH	Flash CO_2 MPH	CO_2 Stripper Overhead MPH	Excess H_2S* MPH	Total MPH
H_2	-	-	24.8	0.1	0.02	24.9
N_2	-	-	0.8	1830.0	-	1830.8
CO	2.50	-	33.9	0.3	0.03	36.73
CO_2	1356.51	194.73	28,331.0	5630.8	62.69	35,575.73
CH_4	3.12	-	197.9	0.5	0.18	201.70
C_2H_4	-	-	84.2	1.3	0.09	85.59
C_2H_6	-	-	112.1	1.7	0.12	113.92
NH_3	-	-	-	0.8	0.10	0.9
H_2S	9.24	8.22	-	0.1	6.85	24.41
COS	-	-	3.1	-	0.03	3.13
CS_2	-	-	-	-	0.03	0.03
Dry	1371.37	202.95	28,787.8	7465.6	70.14	37,897.8
H_2O	112.68	492.37	44.1	24.8	6.91	680.9
Total	1484.05	695.32	28,831.9	7490.4	77.05	37,578.7
Total lb/hr	62,165	17,720	1,257,785	299,649	3628	1,640,947
MM Btu/hr (HHV)	3.73	1.98	134.7	2.30	1.90	144.6

*Return from the citrate process flue gas desulfurization system.

TABLE 9-14. LIQUID FUELS TO INCINERATOR

Compound	Tar	Tar Oil	Phenols	Naphtha	Total
Naphtha ($C_6H_6^+$)	-	-	-	15,622	15,622
Phenols	-	-	11,260	-	11,260
Tar Oils ($\sim C_{16}H_{34}$)	-	48,600	-	-	48,600
Tars ($\sim C_{24}H_{50}$)	88,800	-	-	-	88,800
Total, Lbs/hr	88,800	48,600	11,260	15.622	164,282
MM Btu/hr (HHV)	1,376.4	753.3	156.3	270.0	2,556.0
Nitrogen, Wt%	0.85	0.85	-	-	0.71
Total Nitrogen Lbs/hr	754.8	413.1	-	-	1,167.9
Sulfur, Wt%	0.46	0.26	-	-	0.33
Total Sulfur, Lbs/hr	411.8	125.2	-	-	537.0

TABLE 9-15. INCINERATOR FLUE GAS COMPOSITION

	MPH	MOL % (Wet)
N_2	126,781.8	59.215
O_2	4,277.6	1.998
A	1,601.6	0.748
CO_2	57,726.3	26.962
SO_2	78.6	0.037
SO_3	1.6	0.001
NO_x*	54.75	0.025
Dry	190,522.25	88.986
H_2O	23,581.75	11.014
Total MPH	214,104.00	100.000

Total Lbs/hr	6,724,608
SCFM @ 60°F	1,354,350
ACFM @ 300°F 13 psia	2,237,675

*	% (v)	MPH	Mol %	Lbs/hr
NO	95	52.00	0.024	1,560
NO_2	5	2.75	0.001	126
Total		54.75	0.025	1,686

is equipped with the normal complement of components including firebox, burners, superheater, air heater, economizer, cyclones, and electrostatic precipitators. Estimated efficiency of the units is 86 percent. Combustion Engineering states that special design can accommodate firing of the liquid fuels and coal such that these fuels will pose no combustion problems. They further state that the waste gas streams should be introduced into the upper furnace to avoid interfering with the combustion of liquid fuels and coal.

Particulate Emission--Ash content of the western coal that is used in the Braun study amounts to 7.72 percent of 175,190 lbs/hr of coal feed, or 13,525 lbs/hr. If 80 percent (2, p.IV-9) is assumed to be flyash, with the remaining 20 percent being bottom ash, then 10,820 lbs/hr leave the boiler and enter the cyclone separators. The expected particulate removal efficiency of 75 percent (3, p. 202-205; 4,p.23-20) results in 2705 lbs/hr entering the electrostatic precipitators (ESP's).

The most stringent standards for particulates from combustion are those of New Mexico:

Coal 0.05 lb/MM Btu
Oil 0.005 lb/MM Btu
Gas 0.03 lb/MM Btu

Based on the relative heat inputs of coal, oil, and gas, the allowable emission becomes:

$$E = 0.423(0.05) + 0.546(0.005) + 0.031(0.03)$$
$$= 0.0248 \text{ lb/MM Btu}$$
$$\text{or } 0.0248\ (4678.5) = 116 \text{ lbs/hr}$$

The ESP's need to remove 2,589 lbs/hr. This amounts to a 95.7 percent removal efficiency. These units commonly remove

99 percent or more (3,p.188; 4,p.23-21) of the entering particulate matter. Therefore, no problem is foreseen in meeting or exceeding the most stringent of the present pollution standards. If 99 percent removal is attained, the emission of particulates would be only 27.1 lbs/hr or 0.00578 lb/MM Btu, which is less than 25 percent of the most stringent present standard.

It should be mentioned that additional particulate removal will be accomplished in the spray coolers preceding the SO absorbers in the FGD unit. Therefore, the ultimate particulate emission will be essentially nil for all practical purposes. Conversely, consideration may be given to elimination of the ESP's when FGD is included in the flowsheet.

Flyash from pulverized coal boilers normally ranges in size from 1 to 200 microns, with most of the particles in the size range of 5 to 40 microns (2,p.IV-9, 4, p.23-10). Electrostatic precipitators are effective in removing particles as small as 0.01 micron and as large as 40 to 50 microns (4,p.23-10).

For the low sulfur coal cases, the flyash resistivity will be, at 300 to 350°F, about 3×10^{11} ohm-cm. Therefore, the ESP's must be located upstream of the air heaters where the temperature is around 600°F so that the resistivity becomes sufficienty low (1×10^9 ohm-cm) to sustain efficient particle removal (3,p.188).

CO Emission--The most stringent standard for CO emission from fuel burning equipment is 200 ppm (Illinois). The expected uncontrolled CO emissions from coal, oil and gas fired units are given in the following table (2,p.IV-5):

	Lb/MM Btu	MM Btu/Hr	Lb/Hr
Coal	0.042	1,977.9	83.1
Oil	0.021	2,556.0	53.7
Gas	0.017	144.6	2.4
			139.2

Thus, the total expected CO emission is 139.2 lb/hr or 4.97 MPH. Expressed as ppm by volume this becomes 23.2 ppm (v), or about 10 percent of the most stringent standard.

Hydrocarbon Emission--Illinois has the most stringent standard for hydrocarbon emission from fuel burning equipment. This standard states that incineration must reduce hydrocarbon emission to 10 ppm CH_4 equivalent or less or must convert 85 percent of the hydrocarbons to CO_2 and water. The expected uncontrolled emissions from coal, oil and gas fired units are given in the following table (2,p.IV-5):

	Lb/MM Btu	MM Btu/Hr	Lb/Hr
Coal	0.013	1,977.9	25.7
Oil	0.014	2,556.0	35.8
Gas	0.001	144.6	0.1
			61.6

Thus the expected hydrocarbon emission is 61.6 lbs/hr or 3.84 MPH of CH_4 equivalent, or 17.9 ppm (v). While this figure exceeds the previously mentioned maximum of 10 ppm, the standard is still met by converting the necessary percentage of hydrocarbons to CO_2 and H_2O:

$$\text{HC converted} = (401.2-3.84)(100)/401.2 = 99.0\%$$

This exceeds the required 85 percent conversion of the Illinois standards.

Ammonia Emission--New Mexico's standard for ammonia emission from gasification plants is 25 ppm maximum. Even if none of the ammonia entering the incinerater/boiler is burned, this standard will be met easily. The maximum ammonia concentration in the stack, assuming no combustion, will be 4.2 ppm (v). Since part or all of the ammonia will be burned, the actual ammonia concentration in the stack gases will be considerably less than 4.2 ppm.

Sulfur Emission--Sulfur compounds enter the incinerator/boiler as H_2S, COS, and CS_2 in the gas streams and as organic sulfur compounds in the tar and tar oil. The coal contains pyritic, organic and sulfate sulfur. The sulfur compounds oxidize during incineration and, as determined in Pullman Kellogg's experience, 98 percent are converted to SO_2 and 2 percent to SO_3.

If there were no FGD, the incinerator offgases would contain 71.8 moles of SO_2 and 1.5 moles of SO_3. The emission, therefore, would be 0.98 lb/MM Btu. This emission far exceeds the most stringent standard of 0.2 lb/MM Btu (Okla.) for gas fired equipment, of 0.25 lb/MM Btu (Ky.) for combination fired equipment or of 0.2 lb/MM Btu (Wy.) for coal fired equipment. The FGD step will be described later.

NO_x Emission--The most stringent standard for NO_x emissions is based on New Mexico's regulations which state that maximum emissions of NO_x shall be:

	NO_x, lb/MM Btu
Coal	0.45
Oil	0.30
Gas	0.20

Nitrogen oxides are formed by oxidation of organic nitrogen in the fuel (fuel NO_x) and by oxidation of nitrogen in the combustion air (thermal NO_x).

Typically, about 30 to 60 percent of the fuel nitrogen will be converted to NO_x (3,p.55;6,p.96;7,p.9ff). Therefore, it is clear that reducing the nitrogen content of the fuel will be advantageous. Generally speaking, NO_x formation is a function of temperature (being greater at higher temperatures), oxygen availability and, as mentioned above, the nitrogen content of the fuel. Boiler modifications which lower the flame temperature and reduce the oxygen availability result in lower NO_x formation. These modifications can be:

- Two-stage combustion
- Low excess air firing
- Flue gas recirculation

Expected NO_x emissions for both uncontrolled and controlled boiler conditions are shown below (2,p.III-55; 5,p.21; 6,p.95; 7,p.5):

	NO_x in ppm	
	Uncontrolled	With Boiler Controls
Coal	500	370
Oil	280	150-210
Gas	200	85-110

The allowable NO_x emission, based on New Mexico standards, for the Lurgi low sulfur case becomes:

$$\text{Weighted Average } NO_x = 0.423(0.45) + 0.546(0.3) + 0.031\ (0.2)$$
$$= 0.3604 \text{ lb/MM Btu}$$

Based on a heat input of 4,678.5 MM Btu/hr, the allowable NO_x emission is 1,686.1 lb/hr or 54.75 MPH. Expressed in concentration by volume this becomes 256 ppm(v).

The expected NO_x emission with no boiler controls is calculated as follows:

$$NO_x = .423\ (500) + .546(280) + .031(200)$$
$$= 371\ \text{ppm(v) (which is above the New Mexico standards)}$$

Using similar calculation techniques, the use of boiler controls can be shown to reduce this figure to 241 to 275 ppm.

In the case under study, the nitrogen content of the liquid fuels is 1,168 lb/hr and that of the coal fuel is 1,489 lb/hr for a total of 2,657 lb/hr. With no controls, the expected NO_x emission is projected to be 371 ppm or 79.3 MPH or 2,443 lb/hr of NO_x. In "Processes and Techniques for Control of Nitrogen Oxides" it was shown that about 30 percent of the fuel nitrogen is converted to NO_x or 797 lb/hr of N, equivalent to 1,752 lb/hr of NO_x. Thermal NO_x = 2,443-1,752 = 691 lb/hr.

If 80 percent of the nitrogen in the liquid fuel is removed by hydrodenitrogenation (8,p.14-21,140), then a substantial reduction in NO_x emissions can be expected. The nitrogen content of the fuel oil becomes 234 lb/hr, the coal nitrogen remains unchanged and the total fuel nitrogen becomes 1,723 lb/hr. Conversion of 30 percent of the fuel nitrogen, or 517 lb/hr, to NO_x yields NO_x from fuel of 1,136 lb/hr. The thermal NO_x remains constant at 691 lb/hr and the total NO_x emission becomes 1,827 lb/hr, or 59.3 MPH or 277 ppm(v), a reduction of 25.3 percent from the uncontrolled emissions level.

Therefore, the reduction of the nitrogen content of the liquid fuels is shown to be about as effective in reducing total NO_x emission as are boiler modifications. If both techniques are employed and it is assumed that the percentage reduction of NO_x by boiler modifications is constant at 26 to 35 percent, then the NO_x levels may be reduced to the following:

	NO_x Emission, ppm(v)
Maximum allowed	256
With no control	371
With boiler modifications	241-275
Remove 80% of liquid fuel nitrogen	277
Combined boiler modifications and N removal	180-205

It may be concluded that the most stringent present standards may be met by using boiler modifications and/or liquid fuel denitrogenation techniques. However, if standards are lowered in the future, more complete NO_x removal will be required.

TABLE 9-16, which appeared in "Chemical Engineering" for February 14, 1978, indicates the possible goals for NO_x emissions in 1980 and 1985. The projected 1980 goals may be met by the above mentioned combination of techniques. This is not so for the 1985 goals. The maximum allowable NO_x emissions, based on the relative heat inputs, is projected to be:

$$NO_x\text{, Max. }1985 = 0.423(100) + 0.546(90) + .031(50) = 93 \text{ ppm(v)}$$

This figure is about one half of that attainable with both boiler controls and liquid fuel denitrogenation. Therefore, other methods for NO_x control will be needed if these stringent standards are imposed. There are 48 flue gas denitrogenation processes listed in the literature with 42 being described in detail in a report (7,p.XIV, XV). These processes are generally described

TABLE 9-16. POSSIBLE FUTURE GOALS FOR NO_x EMISSIONS

Source	ppm NO_x at 3% excess O Current technology	 1980 goal	 1985 goal
Utility boilers			
Gas	150 (a)	100	50
Oil	225 (a)	150	90
Coal	550 (a)	200 (c)	100
Industrial boilers			
Gas	150	80	50
Residual Oil	325	125	90
Coal	450	150 (c)	100
Reciprocating engines			
Spark ignition-gas	3,000	1,200 (d)	400
Compression ignition-oil	2,500	1,200 (d)	800
Gas turbines			
Gas	400[150 (b)]	75 (d)	25
Oil	600[225 (b)]	125 (d)	25

(a) Current NSPS.

(b) Estimated achievable with wet control technology.

(c) Developed and field-applied technology.

(d) Developed technology.

Source: EPA Combustion Research Branch (Research Triangle Park, N.C.).

as being dry or wet. There are 5 commercial processes (all dry) which have operated in 50 MW or larger boilers. Of these, the following processes have been chosen for further evaluation:

- UOP/Shell Copper Oxide
- Hitachi Zosen

Both processes use ammonia to reduce nitric oxide to molecular nitrogen as shown below:

$$6\ NO_{(g)} + 4\ NH_{3(g)} \longrightarrow 5N_{2(g)} + 6\ H_2O_{(g)}$$

Each of the processes is capable of NO removal in excess of 90 percent. Therefore, the NO_x emission in the base case without controls could be reduced to about 35 to 40 ppm using these processes.

Processes Producing Phenols, Oils,and Tars: The Alternate Case--

Incineration and steam generation in Lurgi gasification operating with high sulfur coal feed are similar in operation to the base case operation with low sulfur coal but different in the demands on the emission control techniques.

The analysis of the high sulfur coal is given in TABLE 9-12. It is assumed for this illustration of techniques that the required heat input from coal is the same as in the base case. It is also assumed that with the exception of sulfur content the gas and liquid streams would not change significantly in composition and that the flow rates and heating values would be the same as those in the base case. These simplifying assumptions result in the following total heat input to the boilers:

	Gas	Liquid	Coal	Total
Lbs/hr	1,640,947	164,282	159,396	1,964,625
MM Btu/hr (HHV)	144.6	2,556.0	1,977.9	4,678.5
Heat Input, %	3.091	54.633	42.276	100.000

Particulate Emission--The ash content of the eastern coal is 11.3 percent of the 149,832 lb/hr of dry coal feed, or 16,931 lb/hr. Flyash at 80 percent of the total ash amounts to 13,545 lb/hr leaving the boiler. Cyclone removal efficiency of 75 percent results in 3,386 lb/hr entering the ESP's.

Since the relative heat inputs from gas, oil, and coal are the same as in the base case, the allowable particulate emission is the same at 116 lb/hr. The ESP's need to remove 3,270 lb/hr for a removal efficiency of 96.6 percent. As in the base case, no problem is foreseen in meeting or exceeding the most stringent of the present pollution standards. Further, if 99 percent particulate removal is attained in the ESP's, the particulate emission would be 33.9 lb/hr, or 0.00725 lb/MM Btu, which is less than 30 percent of the most stringent present standard.

As previously mentioned, the spray coolers that precede the FGD unit will remove virtually all remaining particulates from the gas stream and therefore the final particulate emission will be essentially zero. And conversely, elimination of the ESP's may be considered when FGD is to be used.

The resistivity of the flyash from high sulfur coal combustion is about 1 x 10^9 ohm-cm at 350°F. This is low enough to allow installation of the ESP's after the air heaters (3,p.188).

CO, Hydrocarbon, and Ammonia Emissions--As in the base case, no problem is seen in meeting present standards.

Sulfur Emission--Sulfur compounds in the incinerator feed are estimated to be, expressed as sulfur:

	With FGD		Without FGD	
	MPH	Lbs/hr	MPH	Lbs/hr
In coal	206.71	6,628	206.71	6,628
In gas	162.74	5,218	117.27	3,760
In liquid	96.00	3,078	96.00	3,078
Total sulfur feed	465.45	14,924	419.98	13,466

If there were no FGD and 90 percent of the sulfur is converted to SO_2, the resulting SO_2 content of the flue gas if 5.64 lbs/MM Btu, far exceeding the most stringent standard. Flue gas desulfurization (FGD) is required. This process will be described later.

NO_x Emissions--With respect to NO_x emissions, the incinerator flue gas is not expected to be significantly different from that of the low sulfur coal, therefore the same arguments are assumed to apply in both cases, leading to the same solutions.

Processes Producing No Phenols, Oils, or Tars--
The Bi-Gas process was selected as being typical of those conversion processes producing no phenols, oils, or tars in the gasifier. In this process the liquid byproduct and waste streams are eliminated and the waste gas streams are reduced drastically. As a result, the steam generator operates as a coal-fed boiler with CO_2 flash gas and CO_2 stripper overhead from the acid gas removal section as the only significant gas streams to be considered.

The required heat input to the boiler is 3,588 MM Btu/hr, supplied by 317,803 lb/hr of dry western coal or 272,025 lb/hr of dry eastern coal.

Particulate Emission--The most stringent standard for particulates is 0.05 lb/MM Btu (N. Mexico). The allowable emission, therefore, is 179.4 lb/hr from the coal-fired boiler. Combustion of the coals is expected to yield 19,627 lb/hr of flyash from western coal and 24,591 lb/hr from eastern coal. Cyclones should reduce the flyash loading of the flue gas to 4,907 lb/hr and 6,148 lb/hr, respectively, for the two coals. ESP's following the cyclones would be required to operate at collection efficiencies of 96.3 percent and 97.1 percent for the two coals, both of which are well within the demonstrated capabilities of ESP's.

As remarked in the discussion of Lurgi operation, the FGD scrubber following the ESP's should reduce particulates to essentially nil, or the FGD scrubber may eliminate the need for ESP's.

CO Emission--The expected uncontrolled CO emission from coal fired boilers is 0.042 lb/MM Btu or 150.7 lb/hr or 5.38 MPH for eastern coal. The CO content of the flue gas is therefore 46.9 ppm(v) and 48.2 ppm(v), respectively, for the two coals, or less than 25 percent of the current most stringent standard (Illinois) of 200 ppm(v).

Hydrocarbon Emission--Standards for hydrocarbon emissions do not apply to fuel burning equipment (boilers). It is noteworthy, however, that the expected hydrocarbon emission from coal fired boilers is 0.013 lb/MM Btu and thus the expected uncontrolled hydrocarbon from western coal combustion is 46.6 lbs/hr or 2.92 MPH of CH_4 equivalent or 25.4 ppm(v). For eastern coal the expected hydrocarbon is 26.1 ppm(v) of CH_4 equivalent.

Sulfur Emission--Sulfur compounds in the boiler feed are estimated to be, expressed as sulfur:

	Western Coal			
	With FGD		Without FGD	
	MPH	Lbs/hr	MPH	Lbs/hr
In coal	65.40	2,097	65.40	2,097
In gas	10.04	322	3.62	116
Total	75.44	2,419	69.02	2,213

	Eastern Coal			
	With FGD		Without FGD	
	MPH	Lbs/hr	MPH	Lbs/hr
In coal	374.99	12,023.5	374.99	12,023.5
In gas	63.01	2,020.4	20.33	652
Total	438.00	14,043.9	395.32	12,675.5

Without FGD and assuming that 98 percent of the sulfur is converted to SO_2, the resulting SO_2 content of the flue gases is 1.21 lb/MM Btu for western coal and 6.92 lb/MM Btu for eastern coal. Both of these values exceed the most stringent standard of 0.2 lbs/MM Btu (Wy.) for coal fired equipment, and FGD is needed.

Liquefaction--
In the SRC II liquefaction conceptual design of Ralph M. Parsons, the fuel for steam and power generation is gas produced in a slagging Bi-Gas gasifier together with tail gas from the sulfur removal system in the process gasifier section.

The boiler is designed with staged combustion and operates with low excess air to minimize NO_x formation. Estimated emission is 100 lbs/hr or 50 ppm(v) of NO_x, which is less than the most stringent standard.

Total SO_2 emissions from the power boilers are estimated at 0.001 lb/ MM Btu, far less than the most stringent standard.

Since the boilers are fired with clean gas, no particulates are expected.

CO emissions are estimated to be 147 ppm with application of the 50 percent excess air correction, less than the 200 ppm (corrected) most stringent standard (Illinois).

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Flue Gas Desulfurization (FGD)

In the preceding discussion of incinerator/boilers in coal conversion plants the necessity for desulfurization of flue gases from gasification process boilers was demonstrated. Since the SRC II process apparently, according to Parsons, produces very little SO_2 as flue gas and no FGD is needed, the discussion that follows will be confined to FGD in gasification.

Consideration of regenerable vs. non-regenerable FGD processes led to the conclusion that production of sulfur for sale was more attractive than scrubbing the flue gases with limestone and disposing of the sludge. Consequently, regenerable processes were studied and the U.S. Bureau of Mines' citrate process was selected for integration into the gasification flowsheets.

With reference to the sulfur balances in Figures 9-47, 9-48, 9-49 and 9-50 the performance of the citrate process FGD scrubber is shown in TABLE 9-17.

For Lurgi gasification, the coal contains about 49 percent of the total sulfur in the feedstock to the incinerator/boiler, whereas for Bi-Gas gasification the coal contains nearly 95 per cent of the total sulfur in the feedstocks to the boiler. Since, in the Lurgi case, the liquid feedstocks have a higher heating value than the coal, the net result is that the SO_2 feed to the scrubber, per million Btu, is about 81.5 percent

TABLE 9-17. FLUE GAS DESULFURIZATION WITH THE CITRATE PROCESS

	Lurgi		Bi-Gas	
	Low Sulfur	High Sulfur	Low Sulfur	High Sulfur
To scrubber*				
as S, lbs/hr	2,571.6	14,924	2,419	14,043.9
as SO_2, lbs/hr	5,138.1	29,818	4,833	28,059.8
as SO_2 lbs/MM Btu	1.10	6.37	1.35	7.82
From scrubber				
as S, lbs/hr	345.4	345.4	359.8	359.8
as SO_2, lbs/hr	690.1	690.1	718.9	718.9
as SO_2, lbs/ MM Btu	0.15	0.15	0.20	0.20
SO_2 removal in scrubber, %	86.57	97.69	85.13	97.44

*Includes the 5 percent excess H_2S that is fed to the scrubber and sent to the incinerator/boiler. See text.

of the SO_2 feed to the Bi-Gas scrubber, and the Lurgi flue gas volume is lower. These effects combine so that the Lurgi FGD system is able to reach a final SO_2 emission level of 0.15 lbs/MM Btu while, at substantially the same sulfur removal efficiency, the Bi-Gas scrubber can reduce the SO_2 emission only to 0.20 lbs/MM Btu. Although both operations meet or exceed the present most stringent emission standard of 0.20 lbs SO_2/MM Btu (Oklahoma), the differences in performance emphasize the variations that may be expected in practice.

In pilot plant operation of the citrate process, SO_2 removal efficiencies of 95 to 98 percent have been obtained. However, for the Bi-Gas high sulfur case to reach the SO_2 emission level of 0.15 lbs/MM Btu, a sulfur removal efficiency of 98.1 percent would be required.

Reference--

Moyes, A. J., Mills, B., and Reeve, R. N., "The Citrex Process for Desulphurisation of Gas Streams." International Conference of European Federation of Chemical Engineers, Salford, England, 6 April 1976. 959*

Claus Sulfur Recovery

In the Lurgi base case, the acid gas stream from the Selexol H_2S removal unit is divided, with about two-thirds being routed to the flue gas desulfurization unit and the remaining one-third being scrubbed with water for ammonia removal before entering the Claus unit. Total gas entering the Claus unit for the Lurgi base case is 780 MPH. This stream contains 9.24 mol percent H_2S (wet basis) and trace amounts of COS and CS_2.

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*Pullman Kellogg Reference File number

In the Claus unit, one-third of the entering H_2S is oxidized with air to SO_2 with liberation of heat:

$$H_2S + 1.5O_2 \longrightarrow SO_2 + H_2O$$

Then the remaining H_2S reacts with SO_2 formed in the oxidation step to form elemental sulfur:

$$2H_2S + SO_2 \longrightarrow (3/x)S_x + 2H_2O$$

This exothermic reaction is carried out over bauxite catalyst. Heat is recovered as steam in the oxidation step as well as after each of the three reaction steps. Overall, the reaction is

$$3H_2S + 1.5O_2 \longrightarrow (3/x)S_x + 3H_2O$$

Molten elemental sulfur formed in the process flows from the sulfur condensers to a sulfur pit. In the Lurgi base case, 26.5 STPD of sulfur are produced in the Claus unit corresponding to a 94.3 percent recovery. In the Lurgi high sulfur coal case, 111.0 STPD are produced corresponding to a 96.4 percent recovery (2)*. It is noteworthy that the Claus plant would approximately triple in size for both cases if a different FGD system (not requiring H_2S) had been chosen.

Tail gas from the final sulfur condenser contains 4,477 ppm(v) sulfur (as S_1) for the low sulfur coal case and 7,891 ppm (v) sulfur for the high sulfur coal case. These concentrations are far too high for discharge to the atmosphere. Therefore, tail gas treatment to reduce sulfur emissions is required. For this treatment the Beavon process was chosen. The Ralph M. Parsons Company, licensors for the Claus/Beavon process, provided design and cost data for the system.

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*Item in reference list following "Beavon Tail Gas Treating Unit"

In the SRC II liquefaction case about 65 percent of the sulfur entering with the feed coal is recovered as salable sulfur in the process gasifier system via the Claus/Beavon process route. About 30 percent of the entering sulfur is recovered as salable sulfur in the fuel gas gasifier system via the Stretford process. (About 1 percent leaves in the ash and 4 percent in the fuel oil product.) In the Claus unit, fed with offgas from the Rectisol acid gas removal system, from the acid gas removal system in the liquefaction section and from sour water stripping, 818 STPD of sulfur are produced, corresponding to a 95 percent recovery. Tail gas from the final sulfur condenser contains 11,972 ppm(v) of sulfur as H_2S and SO_2. As in the Lurgi cases, tail gas treatment is required and for this the Beavon process was selected.

A typical flow sheet for a three-stage Claus process with partial bypass of the acid gas feed is shown in Figure 9-52.

Beavon Tail Gas Treating Unit

The Beavon process, licensed by The Ralph M. Parsons Company, was chosen for treatment of the tail gas from the Claus units in the Lurgi cases and in the SRC II case.

There are about thirty Beavon processes in commercial operation. The Claus/Beavon combination was selected because it appears to be superior to Claus/SCOT or Claus/ARCO processes, at least for these applications, for the following reasons (2):

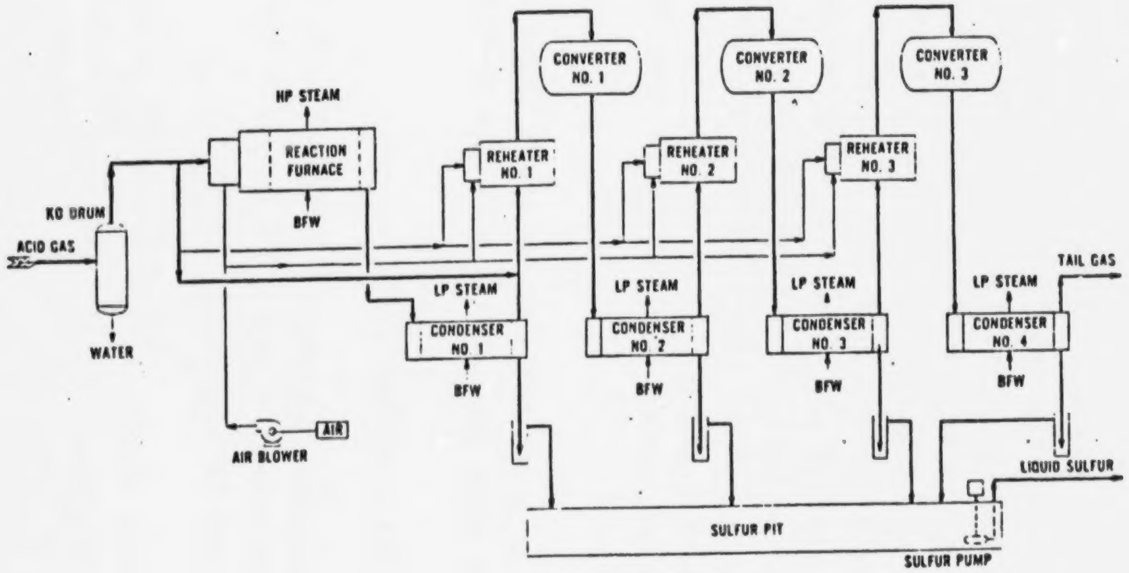
HP STEAM
REACTION FURNACE
BFW
KO DRUM
ACID GAS
WATER
AIR BLOWER
AIR
CONVERTER NO. 1
CONVERTER NO. 2
CONVERTER NO. 3
REHEATER NO. 1
REHEATER NO. 2
REHEATER NO. 3
LP STEAM
CONDENSER NO. 1
BFW
LP STEAM
CONDENSER NO. 2
BFW
LP STEAM
CONDENSER NO. 3
BFW
LP STEAM
CONDENSER NO. 4
BFW
TAIL GAS
LIQUID SULFUR
SULFUR PIT
SULFUR PUMP

Figure 9-52. Claus process flowsheet*

*From Item 1, p. 123, in reference list

- About 15 percent lower capital cost
- Superior operation when processing acid gases containing high CO_2 concentrations
- Lower fuel gas requirements

A simplified flow diagram of the Beavon process is shown in Figure 9-53. The process is based on the catalytic conversion of sulfur species to H_2S by hydrogenation and hydrolysis followed by reaction of H_2S to elemental sulfur in a Stretford unit. Reducing gas is produced via incomplete combustion of hydrocarbon gas (SNG, for example) in air. Tail gas from the Claus unit is mixed with reducing gas and the stream is passed via a catalyst where the following reactions take place at about 700°F (4,p.11):

$$S_8 + 8H_2 \rightarrow 8H_2S$$
$$SO_2 + 3H_2 \rightarrow H_2S + 2H_2O$$
$$COS + H_2O \rightarrow H_2S + CO_2$$
$$CS_2 + 2H_2O \rightarrow 2H_2S + CO_2$$
$$CO + H_2O \rightarrow CO_2 + H_2 \text{ (water-gas shift)}$$

Essentially complete conversion to H_2S is achieved. Heat is recovered in a boiler downstream of the reactor. Final cooling to about 100°F is required before the gas stream enters the Stretford absorber. Here the gas is contacted with an aqueous solution of sodium carbonate activated with sodium metavanadate (SMV) and anthraquinone disulfonic acid (ADA). In the absorber the H_2S reacts as follows:

$$H_2S + Na_2CO_3 \rightarrow NaHS + NaHCO_3$$

In the holding tank following the absorber, SMV reacts with HS^- to form solid elemental sulfur:

$$HS^- + V^{+5} \rightarrow S + V^{+4}$$

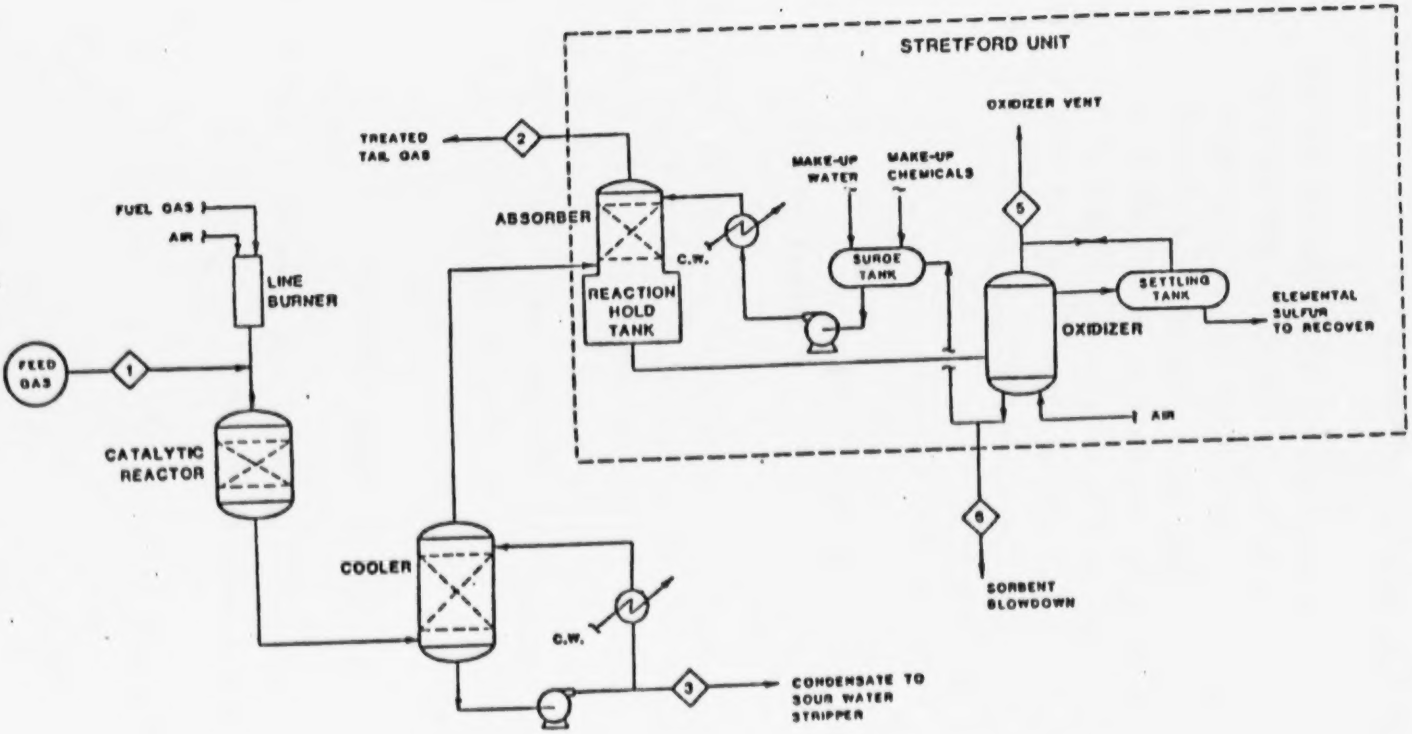
STRETFORD UNIT
TREATED TAIL GAS
2
FUEL GAS
AIR
LINE BURNER
FEED GAS
1
CATALYTIC REACTOR
ABSORBER
REACTION HOLD TANK
C.W.
MAKE-UP WATER
MAKE-UP CHEMICALS
SURGE TANK
OXIDIZER VENT
5
SETTLING TANK
OXIDIZER
ELEMENTAL SULFUR TO RECOVER
AIR
6
SORBENT BLOWDOWN
COOLER
C.W.
3
CONDENSATE TO SOUR WATER STRIPPER

Figure 9-53. Beavon tail gas treating process - typical flow diagram*

*From Item 4 in reference list

This reaction may be shown more completely as:

$$2NaHS + 4\ NaVO_3 + H_2O \longrightarrow 2S + Na_2V_4O_9 + 4NaOH$$

The ADA present in reduced while oxidizing the SMV back to V^{+5}:

$$Na_2V_4O_9 + 2NaOH + H_2O + 2ADA \longrightarrow 4NaVO_3 + 2ADA\ \text{(reduced)}$$

The solution next flows to the oxidizer where oxygen from air is used to oxidize the reduced ADA:

$$2ADA\ \text{(reduced)} + O_2 \longrightarrow 2ADA + 2H_2O$$

Excess air from the oxidizer, free of pollutants, is vented to the atmosphere. The overall reaction for the Stretford process can be written as follows:

$$H_2S + 1/2O_2 \longrightarrow S + H_2O$$

The absorber overhead gas is relatively free of sulfur compounds, containing only 1 ppm H_2S and 62 ppm COS (2).

The sulfur formed appears as a froth containing about 6 to 8 percent sulfur on top of the oxidizer liquid. Filtration and melting with low pressure steam yield a liquid sulfur product of about 99.8 percent purity (4,p.7).

Several side reactions are possible in the process. If the sodium hydrosulfide contacts absorbed oxygen in either the absorber or the oxidizing tank, sodium thiosulfate will form as follows:

$$2NaHS + 2O_2 \longrightarrow Na_2S_2O_3 + H_2O$$

A liquid purge stream is normally needed to rid the system of byproduct salts. This stream is expected to have the following composition:

	Wt %
ADA	0.16
$NaVO_3$	0.69
Na_2CO_3	1.18
Na_2SO_4	7.12
$Na_2S_2O_3$	15.85
H_2O	75.00
	100.00

The quantity to be purged is expected to be 210 and 360 gallons per day, respectively, for Lurgi gasification of low and high sulfur coal (2). By analogy, a purge of about 80 gallons per day is expected for the liquefaction case.

A system has been developed by Parsons for treatment of the purge stream to recover its contained ADA and $NaVO_3$ and reject the byproduct salts as sodium sulfate to eliminate any possible water pollution problem. Alternatively, Union Oil Research has developed a variation of the process which eliminates the formation of byproduct salts and consequently the need for a purge (4,pp.7-8).

Stretford Sulfur Recovery

The Stretford process was selected by Parsons, in their conceptual design for SRC II liquefaction, for recovery of sulfur from the fuel gas gasifier system, apparently because the reactor gases are presumed to contain sulfur only as H_2S. The H_2S is absorbed in a redox solution which is then regenerated in an air blown oxidizer. The sulfur is skimmed from the oxidizer as a froth, melted,and transferred to storage.

Where sulfur is present in the feed gases only as H_2S the Stretford process removal efficiency is nearly 100 percent and yields fuel gas containing about 1 ppm(v) of H_2S. Sulfur production is 370.9 STPD.

References- -

1. Chute, A. E.. "Tailor Sulfur Plants to Unusual Conditions." Hydrocarbon Processing, April 1977.

2. Griebe, M. H., The Ralph M. Parsons Co., Analysis of the Claus/Beavon System Applied to Two Cases for Sulfur Recovery. Private communications, March through May 1978.

3. Pullman Kellogg, Engineering Evaluation of a Process to Produce 250 Billion Btu/Day Pipeline Quality Gas, June, 1972.

4. Beavon, D. K., "Four Years' Experience with the Beavon Sulfur Removal Process ." APCA 70th Annual Meeting, Toronto, June 1977. 905*

5. Moyes, A. J., and Vasan, S., "Holmes-Stretford H_2S Removal Process Proved in Use." Oil and Gas Journal, September 1974. 889*

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*Pullman Kellogg Reference File number

COSTS FOR CONTROL OF GASEOUS EMISSIONS

Development of costs for processes for control of gaseous emissions is based primarily on published data. Updating of the published costs to the end of 1977 was accomplished through use of the Plant Cost Index, compiled and published by "Chemical Engineering" in the May 8, 1978 issue.

The processes selected for investigation are considered to be representative of the best available technology. The developments are simplified hypothetical cases, intended to demonstrate the types of studies that would be required for a more rigorous treatment, and cannot be interpreted or used as definitive estimates. In-depth studies will be needed in order to make specific process recommendations.

Some of the simplifying assumptions that were made for these studies are:

- Power plant size of 500 MW with a heat rate of 9,000 Btu per hour per kilowatt or 4,500 MM Btu per hour
- Costs are calculated to the fourth quarter of 1977 (CE Cost Index = 210, based on 1956 to 1959 = 100) by prorating from the year in which costs are quoted
- Profit and return on investment are not included in the developed costs

More rigorous and extensive studies are needed to consider the many variables that affect the economics of the control technology and, ultimately, the economics of the coal conversion process. Such studies could include the effects on capital and operating costs of:

- Control process size (capacity, throughput)
- Changing analyses of coal feeds

- Conversion plant location
- Conversion plant capacity
- Varying control process efficiency or, stated another way, the effect of meeting varying (increasingly stringent) environmental standards
- Alternate financing schemes for the conversion plant
- Time required for installation of the control process as it affects conversion process downtime for retrofitting or overall construction time for new installations

Particulate Control

The control of particulates in coal conversion facilities will be governed largely by established practices. Particle size, range, density, resistivity, concentration, composition, the degree of removal required, the allowable pressure drop, and other factors will all influence the selection of the particulate control method.

The four most common types of particulate collectors may be arranged in order of increasing efficiency, complexity and cost:

- Cyclone collectors
- Wet scrubbers
- Fabric filters
- Electrostatic precipitators

Cyclones are most commonly used in these applications:

- When particulates are mainly in the coarser size ranges
- When particulate concentrations are fairly high, e.g., above 3 grains per standard cubic foot (SCF)
- When high collection efficiency is not critical
- When they can serve as pre-collectors in conjunction with

other types of collectors that are more efficient in removing fine particulates

Cyclones have the lowest capital cost of the four general types of particulate collectors. Costs of $0.08 to $0.10/ACFM (ACFM = actual cubic feet per minute) for units in the capacity range of 100,000 ACFM have been reported, with special custom designed units costing as much as $0.35/ACFM. Installation will usually add about 25 percent to the cost. These figures are based on 1971 data (1, p.5-10). Escalating these figures to 1977 based on the CE cost index (1971 = 130, 1977 = 210) and adding 15 percent for contractor overhead and profit results in the following installed costs for 100,000 ACFM units:

Costs of Cyclone Collectors, per ACFM

Equipment only, 1971	Installed cost 1977
$0.08	$0.186
0.10	0.232
0.25	0.581
0.35	0.813

High energy wet scrubbers (venturi scrubbers) are normally used where:

- Fine particles must be removed at high efficiency
- Cooling is desired and moisture addition is not objectionable
- Gaseous contaminants as well as particulates are involved
- Volumes are not extremely high (because of the relatively higher operating cost per ACFM)
- Relatively high pressure drop is tolerable
- Contamination of the scrubbing liquid with materials removed from the gas poses no problem

Initial cost (1971) of wet scrubbers sized for about 100,000 ACFM ranged from $0.25 to $0.35/ACFM in carbon steel and about $0.65/ACFM in alloy steel with cost of erection adding about 25 percent (1, p.5-13). Escalating these costs to 1977 and adding contractor overhead and profit yields the following installed costs for 100,000 ACFM units:

Costs of Venturi Scrubbers, per ACFM

Equipment Only, 1971	Installed Cost, 1977
$0.25	$0.580
0.35	0.813
0.50	1.161
0.65	1.509

Fabric filters are normally employed where:

- High efficiencies are desired
- Operation is above the gas dew point
- Temperatures are moderate
- Valuable material is to be collected dry
- Water availability and disposal is a problem

Initial cost of these units (1971) ranged from $0.50 to $1.20 per ACFM for 100,000 ACFM units, depending on the filter medium used. Erection cost is reported to be about 30 percent of the equipment cost (1, p.5-16). Adjusting these costs as before yields the following installed costs for 100,000 ACFM filters:

Costs of Fabric Filters, per ACFM

Equipment Only, 1971	Installed Cost, 1977
$0.50	$1.207
0.75	1.811
1.00	2.414
1.20	2.897

Electrostatic precipitators (ESP's) are most commonly used where:

- Very high efficiencies are required for fine materials
- Volumes of gas are very large
- Water availability and disposal are problems
- Valuable dry material is to be recovered

The purchase price (1971) of an ESP in the 100,000 ACFM capacity range was about $0.80/ACFM while that of one ten times as large was about $0.40/ACFM. Erection cost adds about 70 percent (1,p.5-20). The adjusted installed cost (1977) for 100,000 ACFM becomes $2.527 per ACFM.

The ranges of installed capital costs for the four particulate control devices are shown in Figure 9-54. Variation of capital cost with capacity was obtained by multiplying the base capital cost by the ratio of new to base capacity raised to the 0.7 power.

Operating costs are highly variable, using functions of particulate size, properties, loading, degree of removal required, pressure drop,and other factors. Therefore, no general operating costs have been developed.

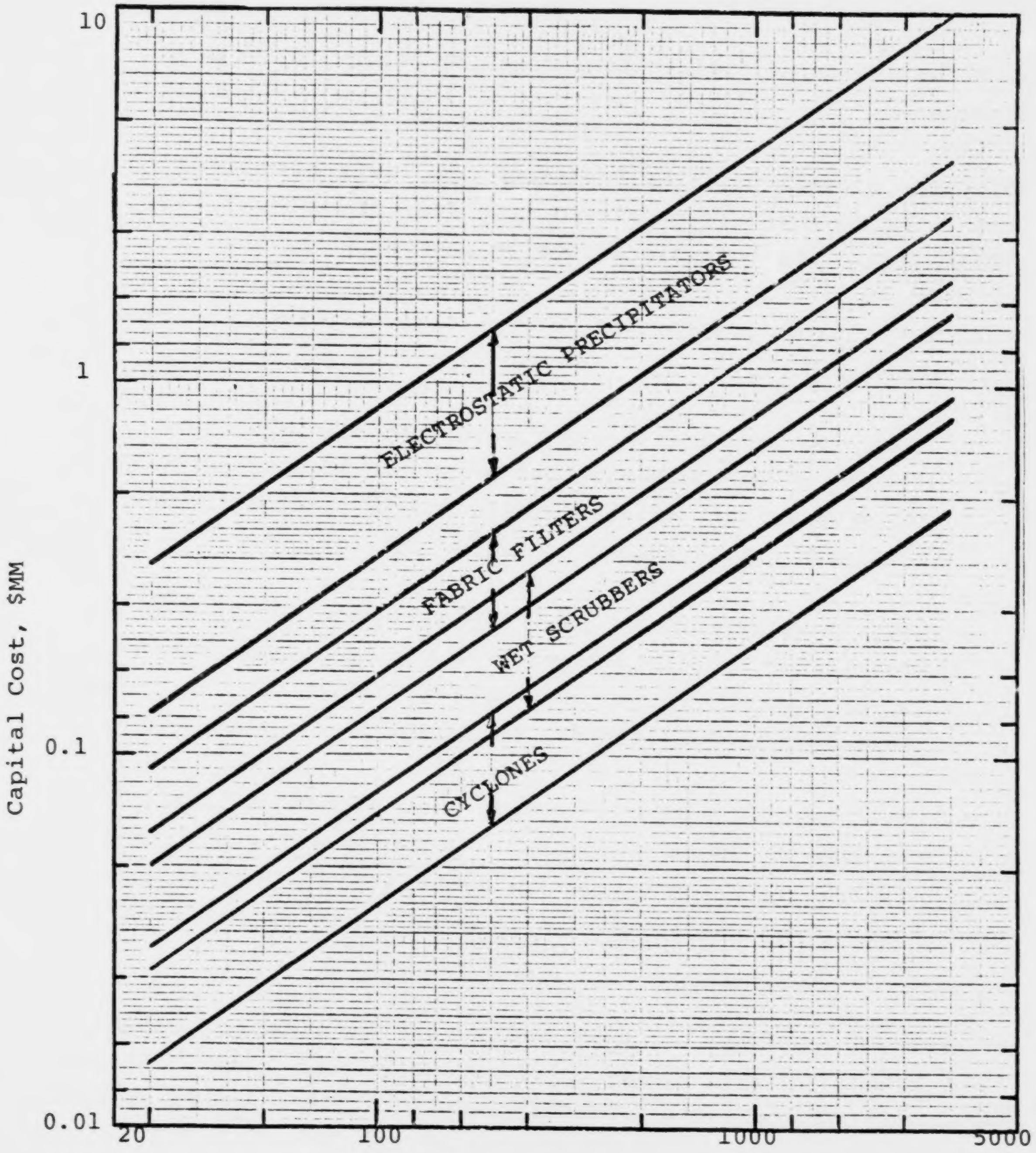

Figure 9-54. Installed capital costs of particulate control devices.

Sulfur Dioxide Control

Desulfurization of Coal by Coal Cleaning and by the Meyers Process--

The range of coal cleaning processes now being practiced in the United States may be generalized into four individual levels of preparation. These levels may be defined as follows:

- o Level 1--no preparation, direct utilization of the run-of-mine product.

  Level 2--removal of gross non-combustible impurities, plus control of particle size and promotion of uniformity (typically 95 percent material yield and 99 percent thermal recovery). Little change in sulfur content.

- o Level 3--single-stage cleaning allowing little component liberation. Particle sizes less than 3/8 inch usually are not prepared. 80 percent material yield and 95 percent thermal recovery. Limited ash and sulfur content.

- o Level 4--multi-stage cleaning with controlled pyrite liberation. Usually incorporated dewatering and thermal drying. 70 percent material yield and 90 percent thermal yield. Maximum ash-sulfur rejection and calorific content of product.

Preparation practice for most coals used by electric utilities lies between levels 2 and 3. The preparation practices for metallurgical coals are typically level 4. The relative costs of these different levels are indicated in TABLE 9-18. The extent to which a specific coal can be cleaned is dependent upon the

TABLE 9-18. PREPARATION PLANT CAPITAL AND OPERATING COSTS (1)(2)

Design Capacity Clean Coal Tons/Yr	Level 4	Level 3	Level 2
Eastern Bituminous Coal			
3,000,000	$25,200,000(3)	$11,200,000	$3,200,000
2,000,000	17,500,000(3)	8,100,000	2,500,000
1,000,000	9,000,000(3)	4,350,000	1,500,000
Western Subbituminous			
Utility Coal (4)			
10,000,000			6,720,000
5,000,000			3,360,000
3,000,000			2,040,000
2,000,000			1,580,000
1,000,000			1,200,000
Cleaning Cost $/Ton (7)	0.80(5) 1.74(6)	0.45(5) 0.87(6)	0.05(5) 0.17(6) 0.05(8) 0.12(8)

(1) From Item 9, pp. 53-56 in reference list

(2) Mid-1974 dollars
Level 4 - Detailed, elaborate facility (75% recovery).
Level 3 - Removal of liberated mineral matter (75% recovery).
Level 2 - Removal of only gross mineral matter (95% recovery).

(3) The capital costs utilized for cleaning eastern bituminous coals at Level 4 ranged between $23,000 and $25,000 per ton of raw feed capacity per hour. Utilizing the "Best Practice" would increase this value to about $30,000 per ton of raw feed capacity per hour. The value would increase to an estimated $40,000 per ton hour if the "Best Cleaning Technology Available" were developed.

(4) Only Level 1 or 2 is applicable. Lignite - Level 1 only considered necessary.

(5) Includes labor, power, maintenance - no amortization or return on investment. Thermal drying adds about 25% to capital costs and 30% to operating costs.

(6) Includes straight line financing at 8% interest, 20 years life and 5% ROI.

(7) Eastern Bituminous coal cleaning - three million tons per year.

(8) Western Subbituminous coal cleaning - ten million tons per year at Level 2.

characteristics of the coal and the sophistication of the preparation process. The limitations are often both economic and technical.

Examples of the operating characteristics for three types of coal cleaning plants are shown in TABLE 9-19. The characteristics of the raw coal feed are different for each plant and represent progressively more difficult separation efforts (2, pp.12, 52). Capital costs for the three plant types were taken from a report by Fluor Utah, Inc. and were updated to 1977. These capital costs are shown in Figure 9-55 as functions of plant feed capacity.

Capital costs were developed for a standard plant size of 75,000 TPD run-of-mine coal delivered to the incoming stockpile for 250 days per year. Included are raw coal receiving and breaking, blending and stockpiling, prepared coal loadout and support facilities.

The yield factor is the TPD output of prepared coal divided by the TPD input of run-of-mine coal. The rated production of a non-standard plant is calculated by dividing the required Btu per day output by the HHV of the prepared coal and by the yield. The cost of a non-standard plant is calculated by multiplying the cost of the standard plant by the ratio of the rated production of the non-standard plant and the rated production of the standard plant raised to an exponent (the cost scaling factor).

Values of these parameters are as follows:

TABLE 9-19. COAL PREPARATION PLANT CHARACTERISTICS

	Raw Coal	Finished Coal	Tons Removed
Hypothetical Crushing and Screening Plant			
Tonnage per day:			
Ash	13,950(18.6%)	13,360(18.40%)	590 (4.2%)
Sulfur	1,130(1.51%)	1,110(1.50%)	20 (1.8%)
Other	59,920	59,780	140 (0.2%)
Total	75,000	74,250(99.0%)	750
Heat Value:			
Btu/lb	10,990	11,030	
Billion Btu/day	1,648	1,638	
Btu yield		99.4%	
Hypothetical Baum Jig Plant			
Tonnage per day:			
Ash	10,350(13.8%)	4,890(7.90%)	5,460(52.8%)
Sulfur	1,065(1.42%)	375(0.61%)	690(64.8%)
Other	63,585	56,610	6,975(11.0%)
Total	75,000	61,875(82.5%)	13,125
Heat value:			
Btu/lb	12,760	13,236	
Billion Btu/day	1,914	1,638	
Btu yield		85.6%	
Hypothetical Heavy Media Plant			
Tonnage per day:			
Ash	10,875(14.5%)	5,015(8.50%)	5,860(53.8%)
Sulfur	1,120(1.49%)	420(0.71%)	700(62.5%)
Other	63,005	53,590	9,415(14.9%)
Total	75,000	59,025(78.7%)	15,975
Heat value:			
Btu/lb	13,084	13,870	
Billion Btu/day	1,963	1,638	
Btu yield		83.4%	

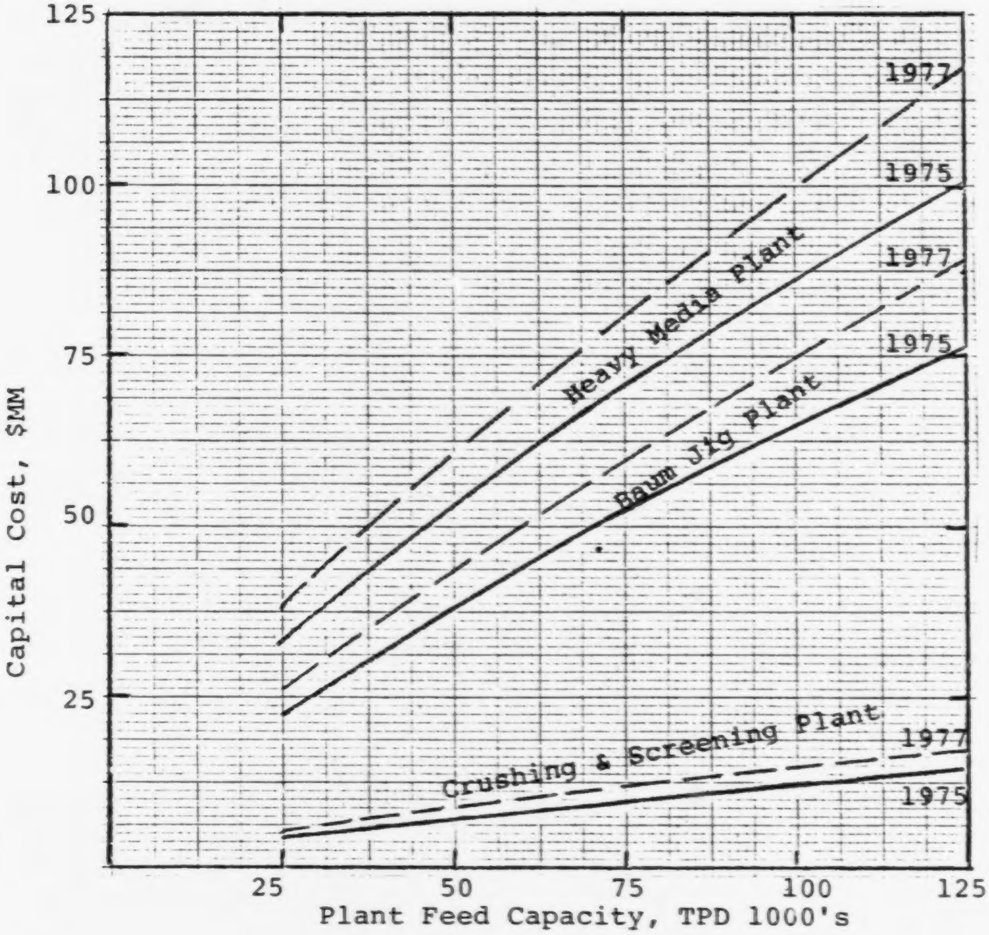

Figure 9-55. Coal Preparation Capital Cost

	Crush & Screen	Baum Jig	Heavy Media
Capital cost (1975), $MM	9.5	52.1	70.2
Standard plant capacity, TPD	75,000	75,000	75,000
Yield factor	0.99	0.825	0.787
Cost scaling factor	0.80	0.75	0.70

The 1975 operating cost for coal preparation in a heavy media plant is stated (9, p.56)(16) to be, per ton of feed coal:

Coal cleaning	$1.90
Coal storage and refuse disposal	0.36
Total operating cost	$2.26

From TABLE 9-19 it is seen that in heavy media cleaning 78.7 percent of the weight of the feed coal is recovered as cleaned product and that the higher heating value (HHV) of the cleaned coal is 6 percent higher than the HHV of the feed coal.

The Meyers process for pyritic sulfur removal was studied by TRW (4, pp.217-230). In this economic study the HHV of the feed coal was taken at 12,291 Btu/lb. The assumption is made for this present comparison that weight recovery and increase in HHV for heavy media cleaning will be the same for feed coal at 12,291 Btu/lb as at 13,084 Btu/lb, thus allowing the two processing methods to be compared on an equal basis.

The operating cost of heavy media cleaning is calculated as follows:

($2.26 x 1,000,000)/(78.7%)(12,291 x1.06)(2000) = $0.110/MM Btu of cleaned coal

The cost of the cleaned coal is calculated from the cost of the feed coal:

Cost of feed coal = \$C/ton
= (C x 1,000,000)/(12,291 x 2,000)
= \$0.0407(C)/MM Btu
Cost of cleaned coal = (0.0407)/(0.787 x 1.06)
= \$0.0488(C)/MM Btu
Incremental cost of cleaned coal = \$0.0081(C)/MM Btu

The total incremental cost of coal cleaning is the sum of the operating cost and the incremental cost of cleaned coal per MM Btu = \$0.110 + \$0.0081(C).

If the incremental cost is to be escalated to 1977, the operating cost portion of the total incremental cost is increased by 17 percent and the total incremental cost per MM Btu becomes \$0.129 + \$0.0081(C).

The Meyers process is an alternate to physical coal cleaning for removal of pyritic sulfur. Pyrite, FeS_2, in the coal is removed by chemical methods to produce ferric sulfate, ferrous sulfate and elemental sulfur. An economic study of the process was performed by TRW, who considered the 4 cases that are outlined in TABLE 9-20 (4, pp.217-230).

The run-of-mine coal fed to each of the four process configuration cases contains 20 percent ash and 3 to 4 percent pyritic sulfur. Approximately 90 to 95 percent of the sulfur is removed during processing. The capital costs include the battery limits and offsite facilities required for processing, interest for construction, startup costs and working capital.

TABLE 9-20. COAL CLEANING WITH THE MEYERS PROCESS*

Case 1:	Cleaned fine coal (14 mesh top size), physical cleaning, Meyers fine coal configuration.
Case 2:	Run-of-mine coarse coal (1/4 inch top size), Meyers coarse configuration.
Case 3 & 4:	Deep cleaned fine and coarse coal with 50% Meyers bypass.

	Case 1	Case 2	Cases 3 & 4
Feed: Rate, TPH	120	100	240
Ash, %	20	20	20
Pyrite S, %	3-4	3-4	3-4
Trillion Btu/yr	23.6	19.7	47.2
Physical Cleaning Ash:			
Rate, TPH	20		40
Ash, %	75		75
Pyritic S, %	10-14		10-14
Trillion Btu/yr	1.4		2.8
Product:			
Rate, TPH	93	85	185-190
Ash, %	6	15	6
Pyritic S, %	0.1	0.2	0.2
Trillion Btu/yr	21.3	17.5	Fine 43.4 Coarse 42.2

*From Item 4, pp. 217-230 in reference list

The required sales prices for processed coal are calculated by the following equation:

$$P = (aX + bY + cZ)/d$$

where

P = Required sales price for processed coal, \$/MM Btu
X = Working capital for raw materials and supplies,\$
Y = Sum of the total plant investment and start-up cost, \$
Z = Annual total operating cost, \$/year
a, b, c = Constants given below
d = Annual energy output, MM Btu/yr

Constant	Investor Financing	Utility Financing
a	0.391	0.140
b	0.505	0.121
c	1.016	1.006

The ugrading costs are determined by deducting the cost of the dirty energy (\$0.41/MM Btu at \$10/ton; \$0.81 at \$20/ton and \$1.22 at \$30/ton).

For Cases 1, 3 and 4 processing costs include ash reduction as well as sulfur reduction. For Case 2, physical cleaning was assumed to be coupled with pyrite removal which results in a major reduction in ash from about 20 percent to about 6 percent.

Prorating cost for plant capacity used the ratio of new plant capacity to base capacity to the 0.7 power.

Capital investment costs for the four cases are shown as a function of capacity in Figure 9-58.

The incremental processing cost for the Meyers process for each of the 4 cases is shown as a function of coal cost, both investor and utility financed, in Figure 9-59. Also shown is the operating cost for heavy media processing.

Inspection of the figures leads to the conclusion that the Meyers process requires more capital than the heavy media process and that its operating costs for the best case are somewhat higher:

	25,000 TPD Feed (1975)	
	Meyers, Case 4	Heavy Media
Capital investment, $MM	101	33
Incremental operating cost*, $/MM Btu	0.330	0.272

*Coal at $20/ton. Utility financed

However, a pyritic sulfur reduction of 95 percent is claimed for the Meyers process while a reduction of only 75 to 85 percent is obtained with heavy media washing. Therefore, in terms of sulfur removal effectiveness, the most favorable Meyers case (Case 4) may be about equivalent to heavy media washing.

Desulfurization of Coal During Combustion in a Fluidized Bed--Fluidized bed combustion (FBC) of coal to generate steam with capture of SO_2 by limestone was compared to a conventional boiler equipped with flue gas desulfurization (FGD) (5). The comparison considered two coals, at 3.6 percent and 0.4 percent sulfur, two capacities, of 100,000 and 400,000 pounds per hour of steam, and three installations, one at a coal fired plant, one at an oil fired plant and one at a grass roots plant. Capital investments

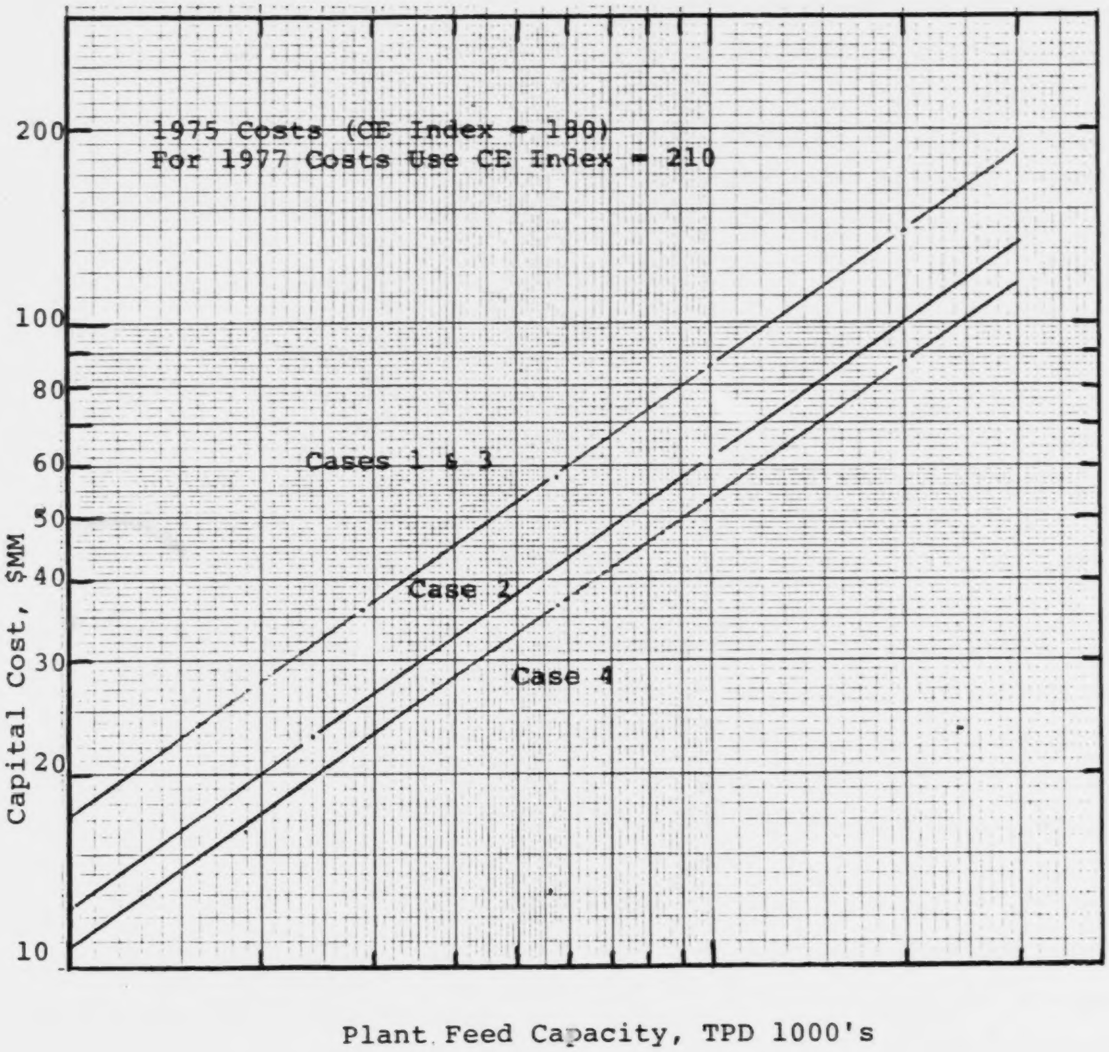

Figure 9-58. Meyers process capital cost.*

*From Item 4 in reference list

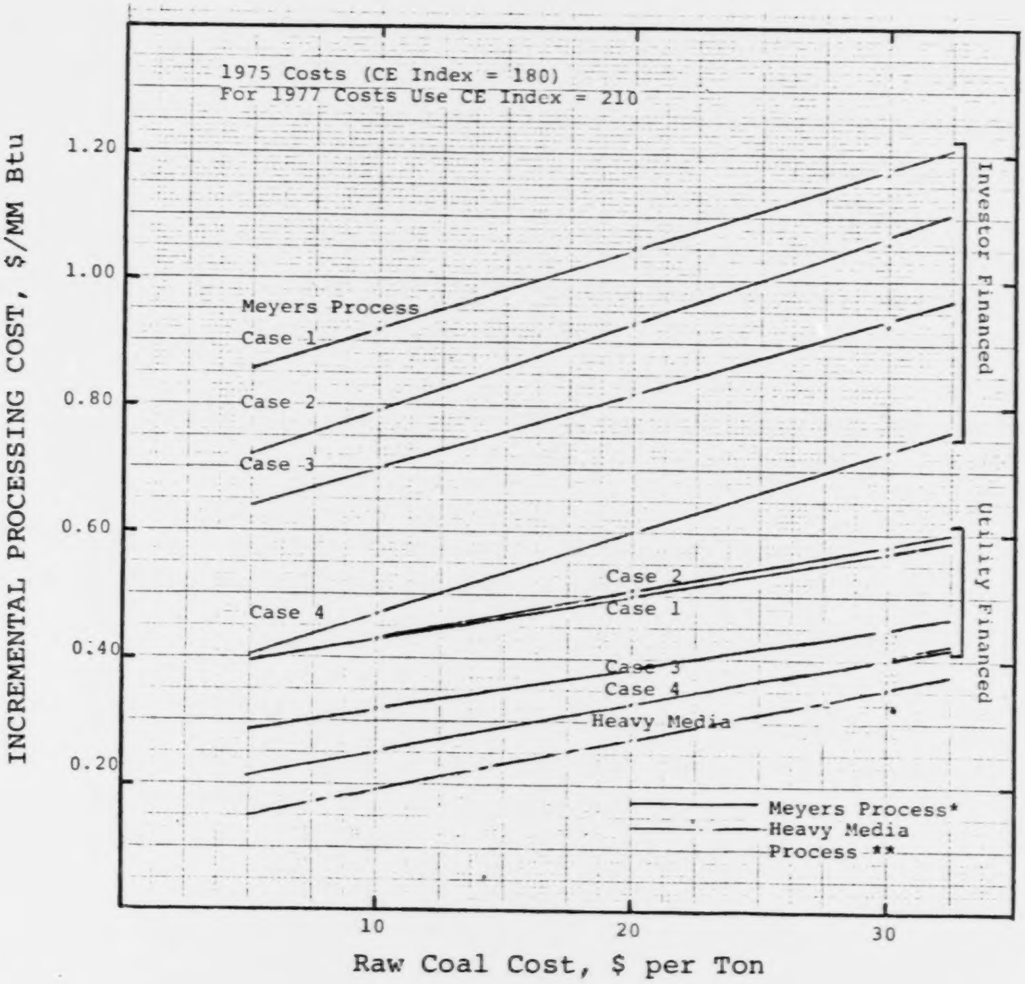

Figure 9-59. Meyers process incremental processing cost.

*From Item 4 in reference list
**From Item 9 in reference list

and operating costs for the cases are shown in TABLE 9-21 for a single boiler added to a coal fired plant, in TABLE 9-22 for a single boiler added to an oil fired plant and in TABLE 9-23 for grass roots boiler plants with backup.

Figure 9-60 compares the capital cost of FBC to conventional installations with FGD in the several configurations and at various steam rates when operating on Illinois No. 6 coal at 3.6 percent sulfur.

Figure 9-61 compares the cost of steam production in the two boiler types in the same configurations and steam rates as in Figure 9-60, when fed with the high sulfur coal.

A review of the report as well as the tables and figures leads to the following conclusions:

- For low sulfur compliance coals, which require no SO_2 controls, a conventional boiler equipped with only an ESP appears to be superior to the FBC.

- For high sulfur coals, the FBC appears to be the better choice. A lower capital investment and steam production cost is obtained for all cases. At a capacity of 400,000 lbs/hr, the FBC shows the following steam cost advantages:

- added to coal fired plant	$0.28/M lb
- added to oil fired plant	$0.34/M lb
- grass roots plant	$0.72/M lb

- One of the advantages of the FBC is its higher heat release rate. It is reported to be 100,000 Btu/hr/cubic foot of expanded bed volume (50,000 to 60,000 Btu/hr/CF of firebox) as opposed to 20,000 Btu/hr/CF of firebox for a conventional boiler (5, p.11).

TABLE 9-21. COMPARISON OF INVESTMENTS AND COST OF STEAM (EX FUEL) FOR SINGLE BOILER ADDED TO COAL FIRED PLANT*

Fuel (1)	High Sulfur Coal				Low Sulfur Coal (2)			
Boiler Type	Fluidized Bed Combustion		Conventional With Scrubber		Fluidized Bed Combustion		Conventional With ESP	
Steam Rate, KPPH	100	400	100	400	100	400	100	400
Investment, M$								
Fuel Handling Additions	0.6	0.9	0.2	0.3	0.6	0.9	0.2	0.3
Boiler and Stack	3.1	7.6	2.9	8.6	3.1	7.6	2.9	8.6
Envtl. and Waste Disp.	1.1	2.9	2.3	6.1	0.9	2.4	0.7	1.8
Total, M$	4.8	11.4	5.4	15.0	4.6	10.9	3.8	10.7
Unit Cost of Steam (ex Fuel), ¢/k lb.								
Direct Op. Costs (ex Fuel and BFW)	125	95	131	100	83	53	64	45
Boiler Feed Water	60	60	60	60	60	60	60	60
Capital Charges	122	72	137	95	117	69	96	68
Total, ¢/k lb. (ex fuel)	307	227	328	255	260	182	220	173

(1) The same type fuel is assumed to be fired in the existing boilers as in the new boiler, i.e., no cases are considered in which low sulfur coal and high sulfur coal are fired simultaneously to different boilers of the same plant.

(2) Low sulfur coal by definition for these cases is assumed to be suficiently low in sulfur so that no SO_2 controls are needed to meet whatever environmental limits are applicable.

*From Item 5 in reference list.

TABLE 9-22. COMPARISON OF INVESTMENTS AND COST OF STEAM (EX FUEL) FOR SINGLE BOILER ADDED TO OIL FIRED PLANT

Fuel	High Sulfur Coal				Low Sulfur Coal (1)				Low Sulfur Fuel Oil(1)	
Boiler Type	Fluidized Bed Combustion		Conventional With Scrubber		Fluidized Bed Combustion		Conventional With ESP		Package	
Steam Rate, KPPH	100	400	100	400	100	400	100	400	100	400
Investment, M$										
Fuel Handling Allowance (2)	1.8	2.7	1.8	2.7	1.9	2.9	1.9	2.9	0.1	0.2
Boiler and Stack	3.1	7.6	2.9	8.6	3.1	7.6	2.9	8.6	1.5	3.7
Envtl. and Waste Disp.	1.3	3.4	2.6	6.9	1.1	2.9	1.0	2.6	--	--
Total, M$	6.2	13.7	7.3	18.2	6.1	13.4	5.8	14.1	1.6	3.9
Unit Cost of Steam (ex Fuel), ¢/k lb.										
Direct Op. Costs (ex Fuel and BFW)	142	102	150	108	100	60	83	52	46	31
Boiler Feed Water	60	60	60	60	60	60	60	60	60	60
Capital Charges	157	87	185	115	155	85	147	89	41	25
Total, ¢/k lb. (ex fuel)	359	249	395	283	315	205	290	201	147	116

(1) Low sulfur coal and fuel oil by definition are sufficiently low in sulfur that no SO_2 controls are needed to meet whatever environmental limits are applicable.

(2) In some cases where coal is reliably available by truck delivery, the capital costs for fuel receipt and storage could be significantly reduced. In such cases, however, the delivered price of coal would rise more or less correspondingly so that the overall cost of steam would not be changed markedly.

TABLE 9-23. COMPARISON OF INVESTMENTS AND COST OF STEAM (EX FUEL) FOR GRASS ROOTS BOILER PLANTS WITH BACKUP

Fuel	High Sulfur Coal				Low Sulfur Coal (1)				Low Sulfur Fuel Oil(1)	
Boiler Type	Fluidized Bed Combustion		Conventional With Scrubber		Fluidized Bed Combustion		Conventional With ESP		Package	
Steam Rate, KPPH	100	400	100	400	100	400	100	400	100	400
Investment, M$										
Fuel Handling Allowance (2)	1.8	2.7	1.8	2.7	1.9	2.9	1.9	2.9	0.6	1.4
Boiler and Stack	5.8	14.3	5.4	16.0	5.8	14.3	5.4	16.0	2.6	6.4
Envtl. and Waste Disp.	1.9	5.0	4.6	12.2	1.7	4.5	1.6	4.2	--	--
Total, M$	9.5	22.0	11.8	30.9	9.4	21.7	8.9	23.1	3.2	7.8
Unit Cost of Steam (ex Fuel), ¢/k lb.										
Direct Op. Costs (ex Fuel and BFW)	180	119	206	135	139	79	121	71	77	44
Boiler Feed Water	60	60	60	60	60	60	60	60	60	60
Capital Charges	241	140	299	196	238	137	226	147	81	49
Total, ¢/k lb. (ex fuel)	431	319	565	391	437	276	407	278	218	153

(1) Low sulfur coal and fuel oil by definition are sufficiently low in sulfur that no SO_2 controls are needed to meet whatever environmental limits are applicable.

(2) In some cases where coal is reliably available by truck delivery, the capital costs for fuel receipt and storage could be significantly reduced. In such cases, however, the delivered price of coal would rise more or less correspondingly so that the overall cost of steam would not be changed markedly.

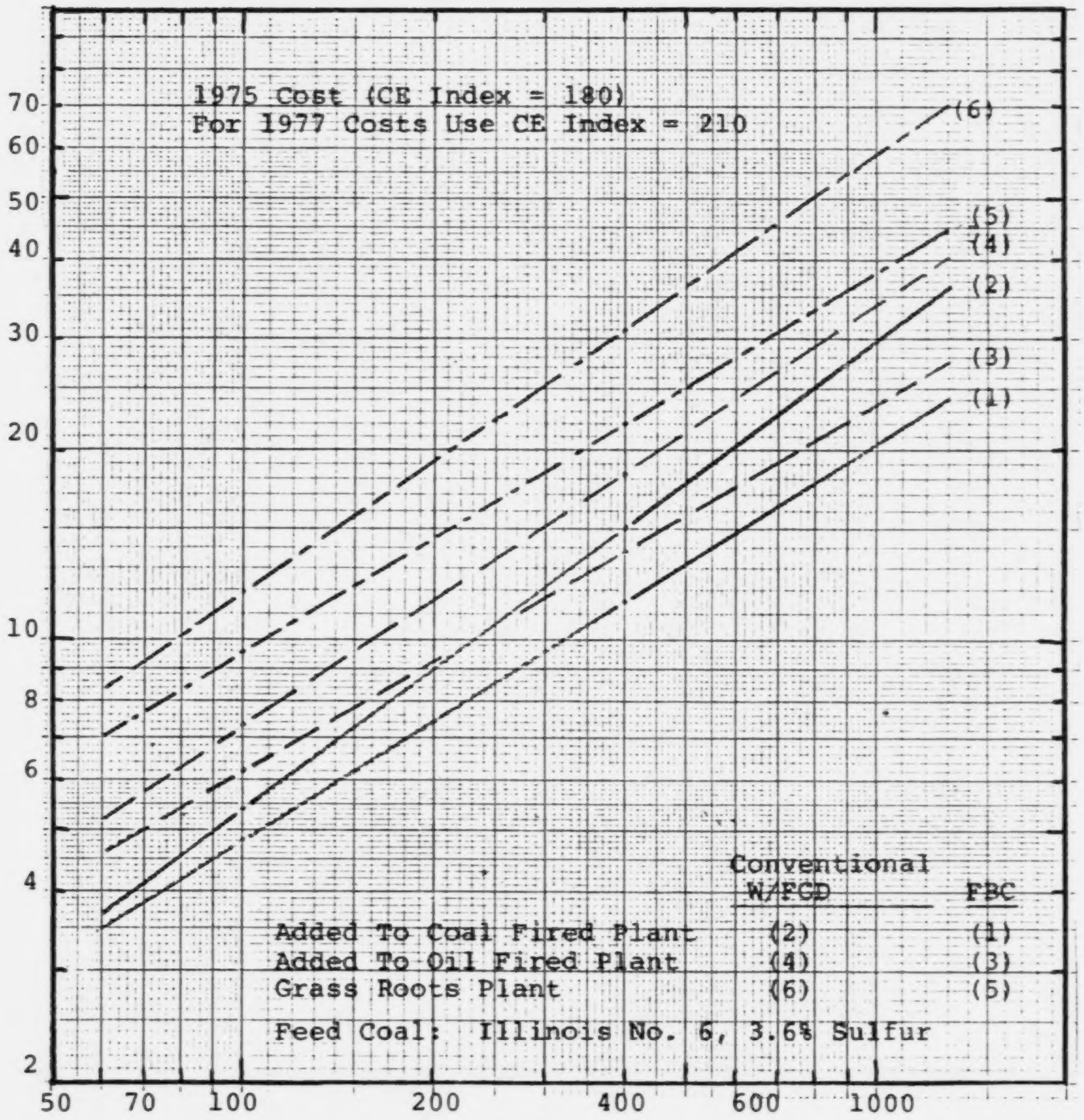

Figure 9-60. Capital cost comparison: FBC vs. conventional boiler with FGD*

*From Item 5 in reference list

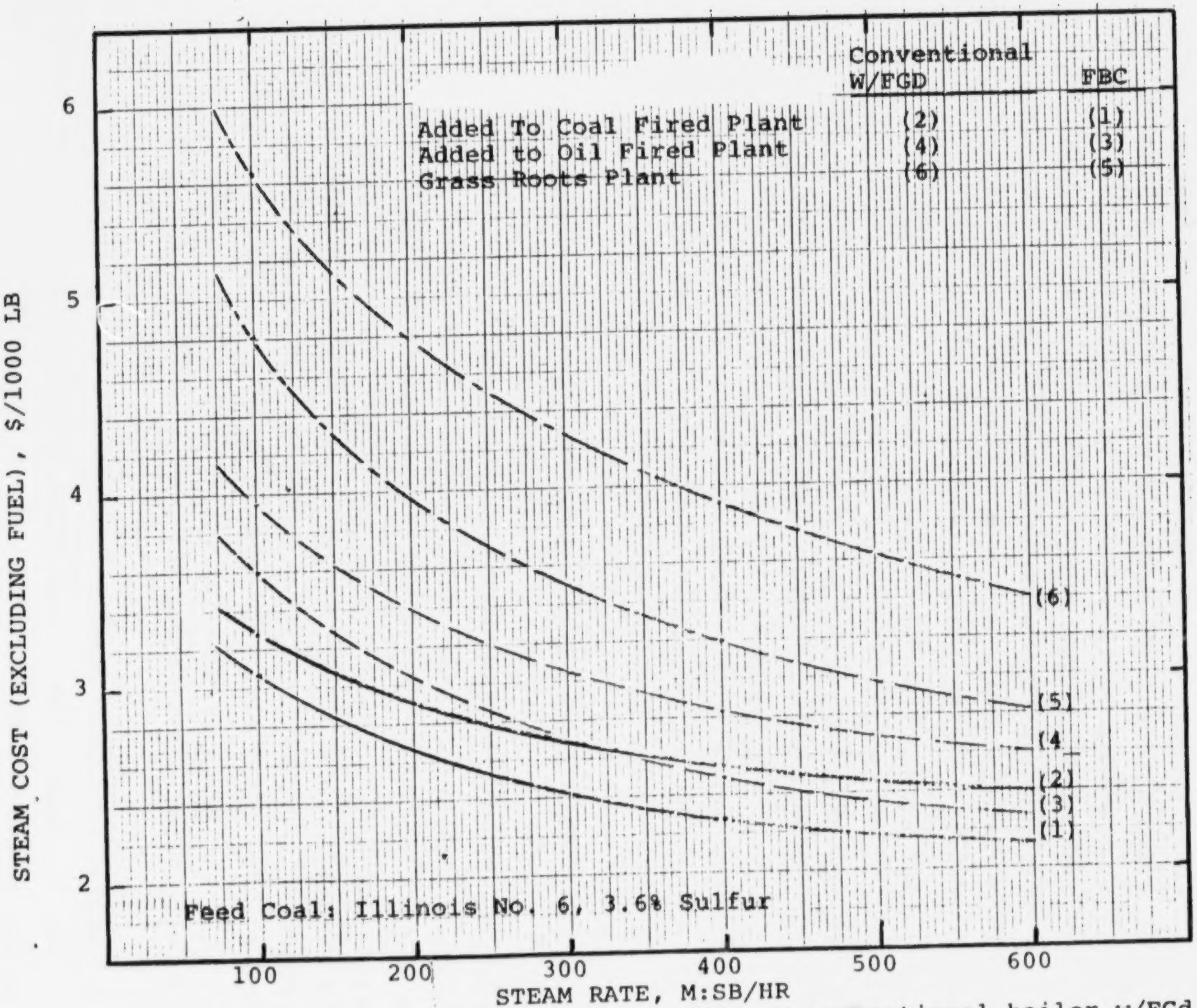

Figure 9-61. Steam cost comparison: FBC vs. conventional boiler w/FGd.*
*From Item 5 in reference list

- Another advantage of the FBC is the formation of $CaSO_4$ which is disposed of in reduced quantities as a dry solid mixed with ash and unreacted limestone. Wet limestone scrubbing systems form a sludge consisting of $CaSO_3 \cdot 1/2H_2O$, $CaSO_4 \cdot 2H_2O$, unreacted limestone, flyash and water (normally about 50%). Assuming a Ca/S mol ratio of 3 for the FBC and 1.2 for the wet limestone scrubbing system results in a 20 to 30 percent reduction in solid wastes for the FBC (5,pp.126-130).

- Thermal NO_x emissions should be reduced for the FBC system due to its lower operating temperature of 1,400° to 1,600°F as opposed to 2,500°F or higher for conventional boilers (5,p.9). Exxon estimates NO_x at 0.5 lb/MM Btu for FBC as compared to 0.7 for a conventional boiler (5,p.128).

- Further development of the FBC appears warranted.

Flue Gas Desulfurization--

Desulfurization of stack gases by either dry or wet gas cleanup systems has been studied for several years. There are large scale test programs presently underway to determine system reliability and cost for several sulfur dioxide removal processes. Some of the EPA-sponsored tests are summarized in TABLE 9-24. Cost data for these processes have been published (12)(13) in great detail and may be referred to for support of the summarizations that follow.

The design of sulfur dioxide removal facilities is dependent upon the actual quantities of gas to be handled and the sulfur dioxide emission rates. TABLE 9-25 shows these parameters for several variations.

TABLE 9-24. EPA-SPONSORED STACK GAS DESULFURIZATION DEMONSTRATION SYSTEMS*

EPA-Sponsored Process (byproduct)	Cooperating Utility	Process Developer	Location	Unit Size and Type
Limestone slurry scrubbing (sludge)	TVA	Bechtel and others	Shawnee Unit 10 Paducah, Ky.	10 MW coal
Lime slurry scrubbing (sludge)	TVA	Chemico, Bechtel and others	Shawnee Unit 10 Paducah, Ky.	10 MW coal
Magnesia slurry scrubbing-regeneration (98% sulfuric acid)	Boston Edison	Chemico-Basic	Mystic Station 6 Boston, Mass.	155 MW oil
Catalytic oxidation (reheat) (80% sulfuric acid)	Illinois Power	Monsanto	Wood River Station 4 East Alton, Ill.	110 MW coal
Sodium scrubbing-regeneration (sulfur)	Northern Indiana Public Service Co.	Davy Powergas Allied Chemical	D. H. Mitchell Station 11 Gary, Ind.	115 MW coal

*From Item 12 in reference list

TABLE 9-25. FLUE GAS AND SULFUR DIOXIDE EMISSION RATES FOR NEW COAL FIRED POWER PLANTS

Plant Size, MW	Fuel Sulfur Content, %	Gas Flow to FGD, M ACFM (310°F)	SO_2 to FGD lbs/hr
200	3.5	630	9,310
500	2.0	1,540	13,010
500	3.5	1.540	22,760
500	5.0	1,540	32,510
1,000	3.5	2,980	44,000

TABLE 9-26. REQUIRED REMOVAL EFFICIENCIES IN FGD UNITS

Sulfur Content of Fuel, %	Particulate Removal, %	SO_2 Removal, %
2.0	98.7	58.9
3.5	98.7	76.3
5.0	98.7	83.4

The FGD systems are designed to meet the existing federal allowable emissions for coal fired boilers of 0.1 lb/MM Btu heat input for particulates and 1.2 lbs/MM Btu heat input for sulfur dioxide. TABLE 9-26 shows the required removal efficiencies to meet the standards.

In TABLE 9-27 are shown the estimated capital investments for FGD units using the limestone slurry and lime slurry processes, both of which produce sludge that requires disposal, and the magnesia slurry and catalytic oxidation processes, both of which produce sulfuric acid. The comparison is based on the following:

- New coal fired units
- Midwest location
- Costs as of 1974
- 90 percent sulfur dioxide removal
- 30 years life
- Stack gas reheat to 175°F
- On site disposal pond 1 mile from power plant
- No flyash disposal facilities

TABLE 9-28 shows the operating costs for the FGD units under consideration, based on 7,000 hours of operation per year and 1975 operating costs.

Figure 9-62 shows the effect of power unit size on the capital investment for the limestone slurry and lime slurry FGD processes. The same presentation is made in Figure 9-63 for the magnesia slurry and catalytic oxidation processes.

Figure 9-64 illustrates the effect of the sulfur content of the coal feed on the operating costs for the 500 MW unit. Figure 9-65 shows the effect of plant size on the operating costs with fuel containing 3.5 percent sulfur.

TABLE 9-27. CAPITAL INVESTMENT FOR FGD UNITS

Plant Capacity, MW	200	500			1000
Sulfur in Fuel, %	3.5	2.0	3.5	5.0	3.5
Limestone Slurry, $MM	13.03	22.60	25.16	27.34	37.73
$/KW	65.16	45.20	50.32	54.69	37.73
Lime Slurry, $MM	11.75	20.23	22.42	24.27	32.77
$/KW	58.75	40.46	44.84	48.54	32.77
Magnesia Slurry, $MM	14.14	22.96	26.41	29.36	38.87
$/KW	70.60	45.92	52.82	58.71	38.87
Catalytic Oxidation,					
$MM	19.54	42.52	42.74	42.93	69.89
$/KW	97.69	85.04	85.47	85.86	69.89

TABLE 9-28. OPERATING COST FOR FGD UNITS

Plant Capacity, MW	200	500			1000
Sulfur in Fuel, %	3.5	2.0	3.5	5.0	3.5
Limestone Slurry					
Mills/KWH	2.80	1.94	2.20	2.43	1.70
¢/MM BTU*	30.43	21.55	24.44	27.00	19.54
Lime Slurry,					
Mills/KWH	2.97	1.98	2.31	2.62	1.79
¢/MM BTU*	32.28	22.00	25.67	29.11	20.57
Magnesia Slurry,					
Mills/KWH	3.41	2.15	2.63	3.08	2.05
¢/MM Btu*	37.07	23.88	29.22	34.22	23.56
Catalytic Oxidation					
Mills/KWH	3.02	2.51	2.54	2.55	1.99
¢/MM Btu*	32.83	27.88	28.22	28.33	22.87

*HHV heat input of fuel

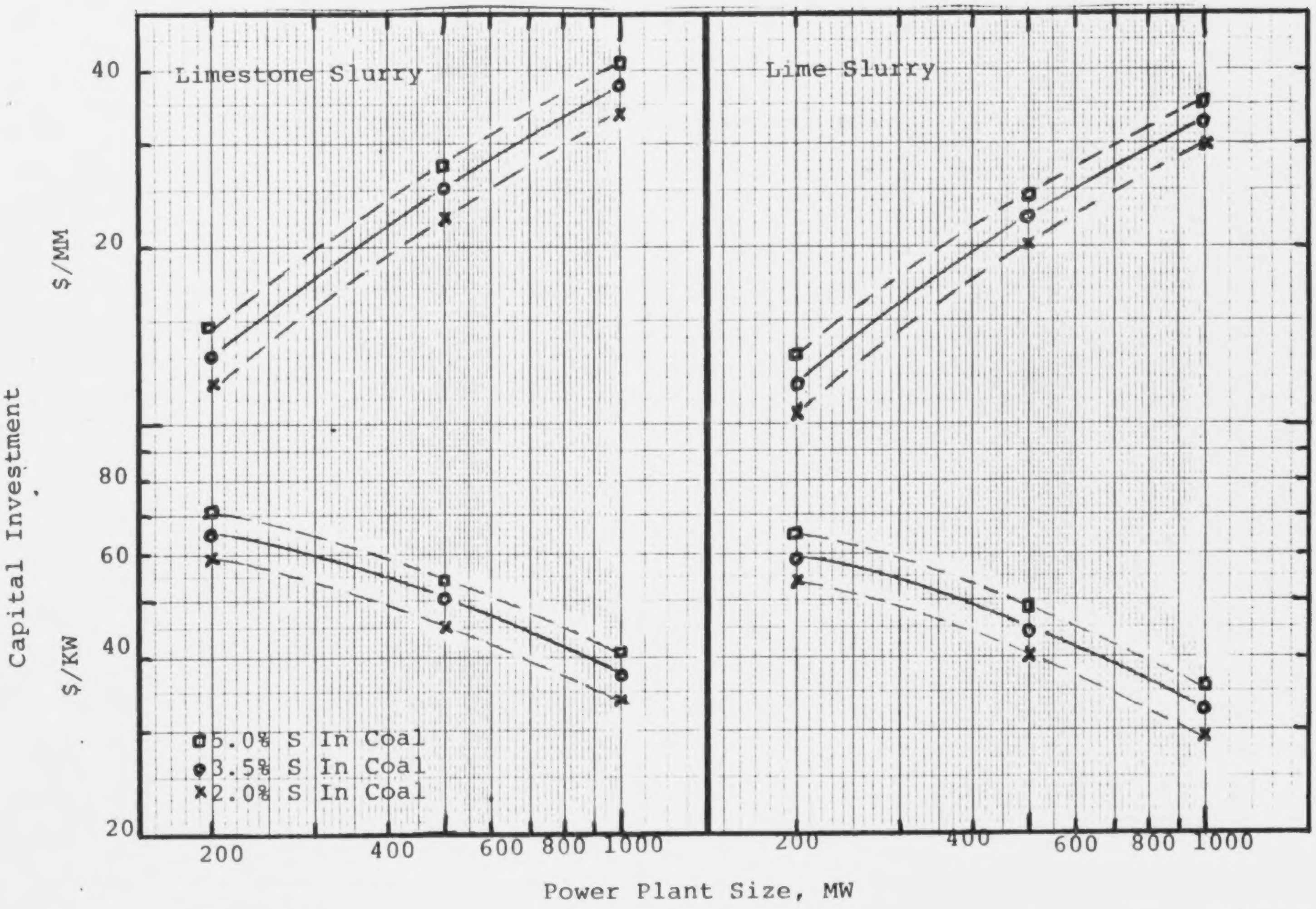

Figure 9-62. Capital investment for limestone and lime slurry FGD processes.

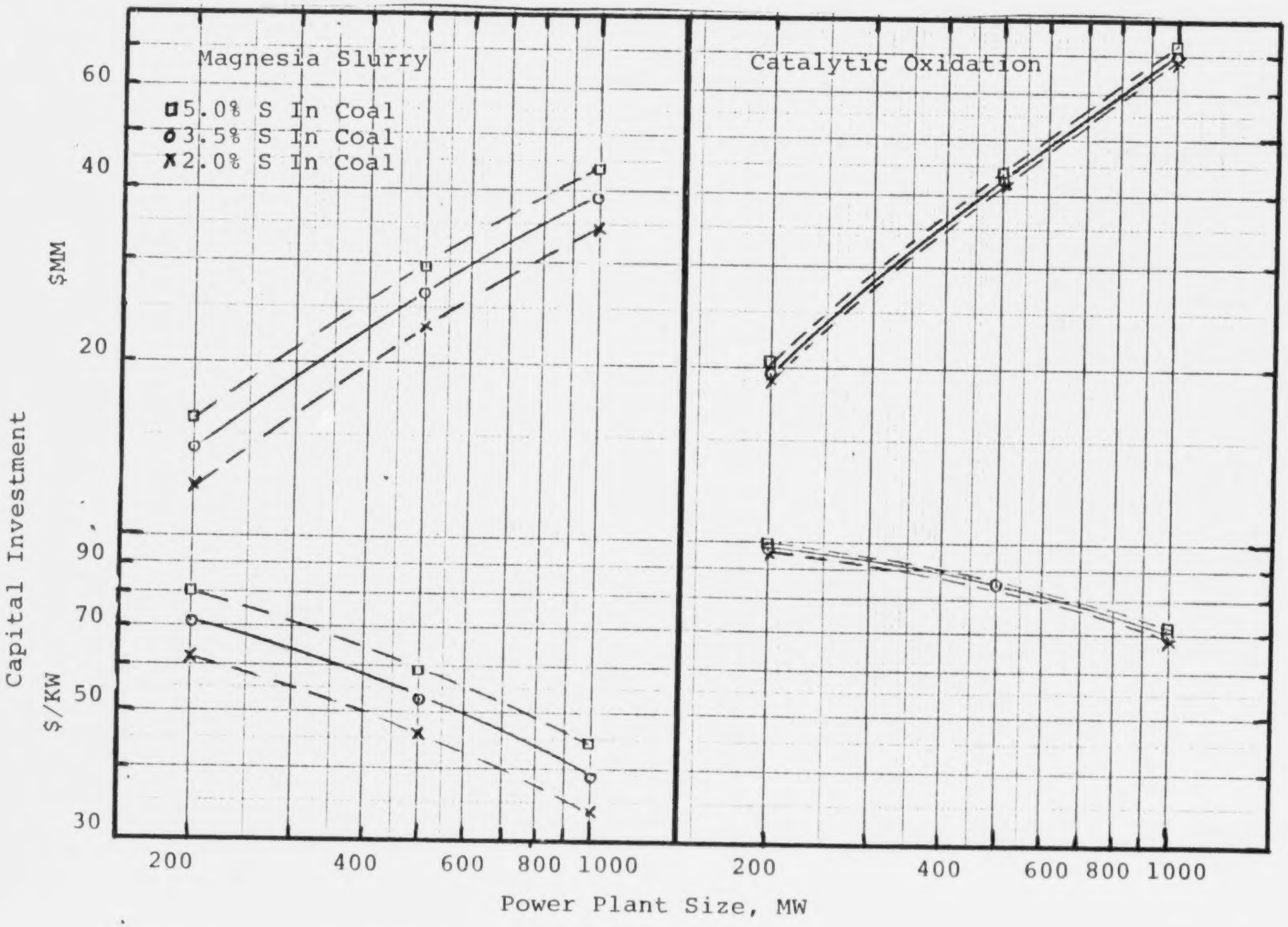

Figure 9-63. Capital investment for magnesia slurry and catalytic oxidation FGD processes.

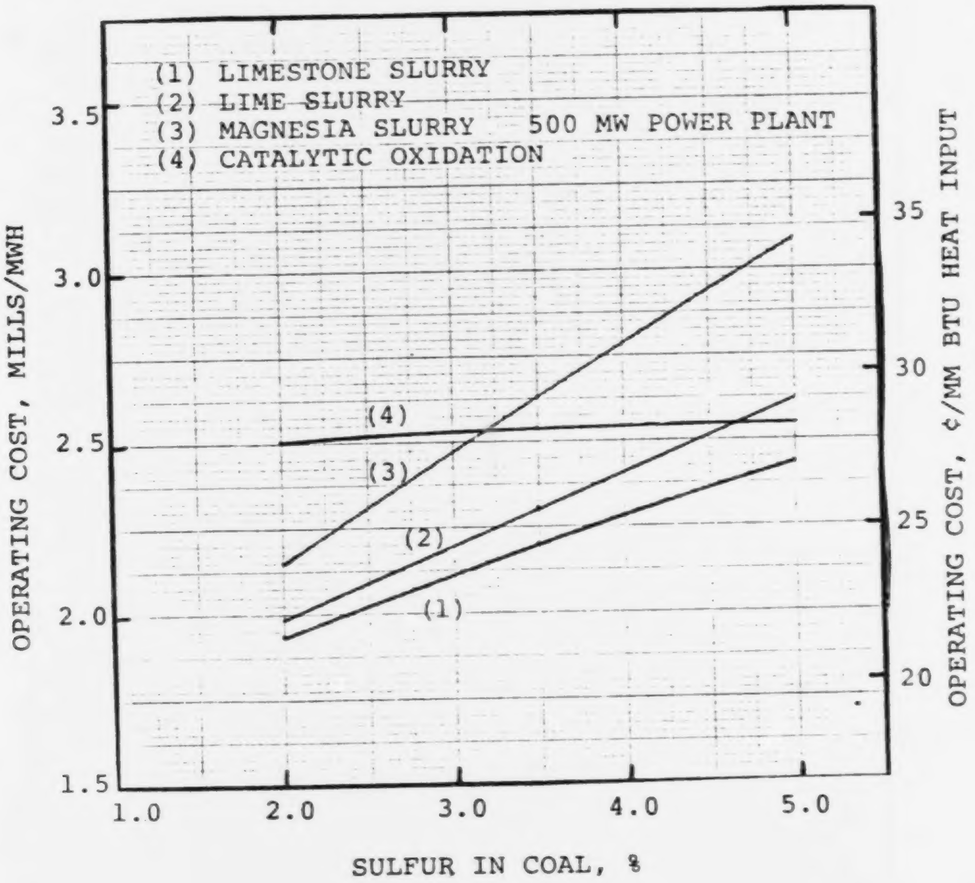

Figure 9-64. Effect of sulfur content of coal feed on FGD operating cost.

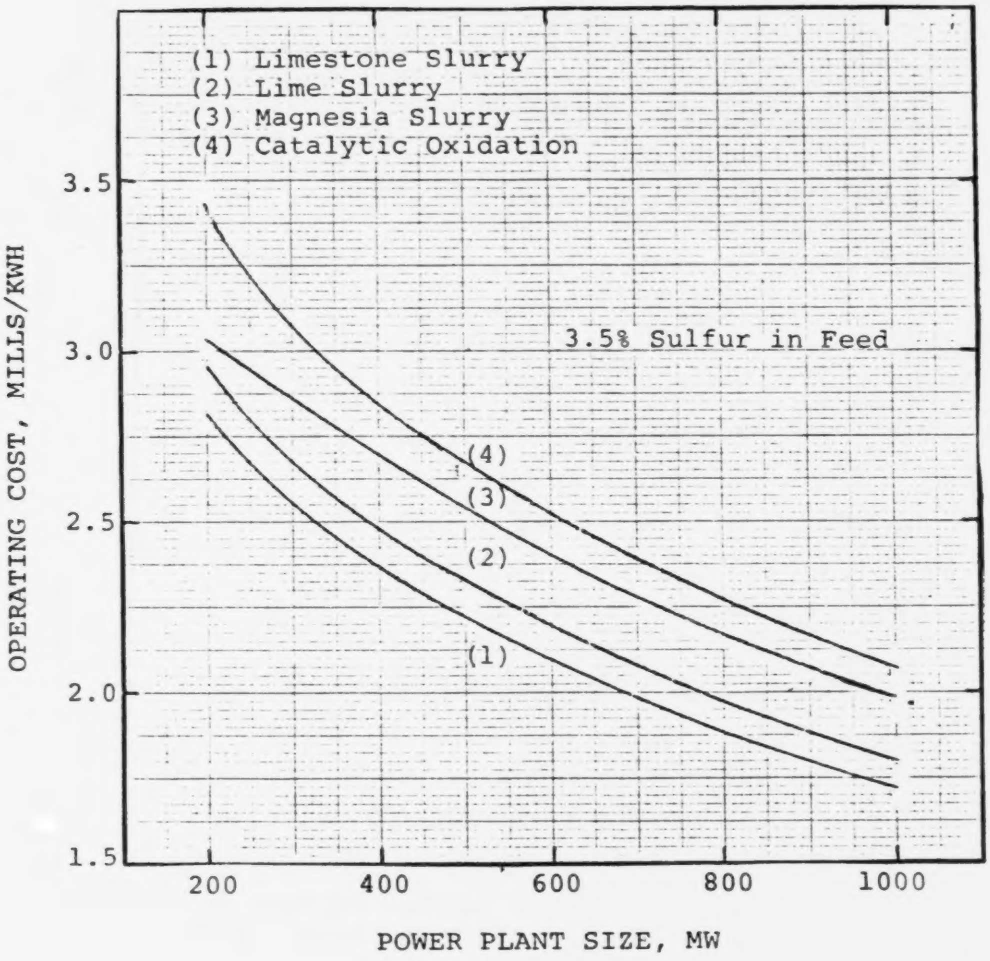

Figure 9-65. Effect of plant size on FGD operating costs.

Comparison of these data show that capital investment for the throwaway processes is lower than for the product recovery process and that the lime slurry process has the lowest capital requirement. Operating costs for the limestone slurry process are, however, lower than those for lime slurry.

It is of interest to note in Figure 9-64 that variation in the sulfur content of the feed coal has very little effect on the operating costs of the catalytic oxidation process.

Capital investment and operating costs of the limestone slurry process and the citrate process for sulfur recovery are compared in TABLE 9-29 for a 500 MW power plant in a midwest location fed with coal containing 3.5 percent sulfur and for removal of 90 percent of the sulfur dioxide from the flue gas. The citrate process costs were calculated from published data and information (13) to a basis comparable to that of the limestone slurry process in order to develop the comparison. The hydrogen sulfide that is needed for the citrate process was assumed to be supplied from the feed to the sulfur recovery section of the coal conversion process.

The capital and operating costs are shown graphically in Figures 9-66 and 9-67, respectively, for the processes by years. It is of interest to note the apparent convergence in both figures to an intersection around 1984.

Comparison of the costs for the citrate process and the limestone slurry process indicates that the citrate process with credit for sulfur sales is a close competitor of the limestone slurry process. Further, the citrate process has the advantage of sulfur production as opposed to sludge production. For these reasons the citrate process was selected for inclusion in further economic studies.

TABLE 9-29. COST COMPARISION OF LIMESTONE SLURRY AND CITRATE FGD

	Limestone Slurry	Citrate (1)
Capital Investment, $MM		
1974	25.16	-
1977	-	47.63
1979	43.32	51.35
Operating Cost, $MM/yr		
1975	7.70	-
1977 (2)	-	12.28
1977 (3)	-	13.67
1980 (2)	-	14.40
1980 (3)	13.56	15.35
Operating Cost, Mills/KWH		
1975	2.20	-
1977 (2)	-	3.51
1977 (3)	-	3.91
1980 (2)	-	4.11
1980 (3)	3.87	4.39
Operating Cost, ¢/MM Btu (4)		
1975	24.45	-
1977 (2)	-	39.00
1977 (3)	-	43.41
1980 (2)	-	45.70
1980 (3)	43.03	48.72

(1) Hydrogen sulfide supplied by coal conversion process (2) With sulfur sales at $40 per ton (3) without sulfur sales (4) HHV heat input

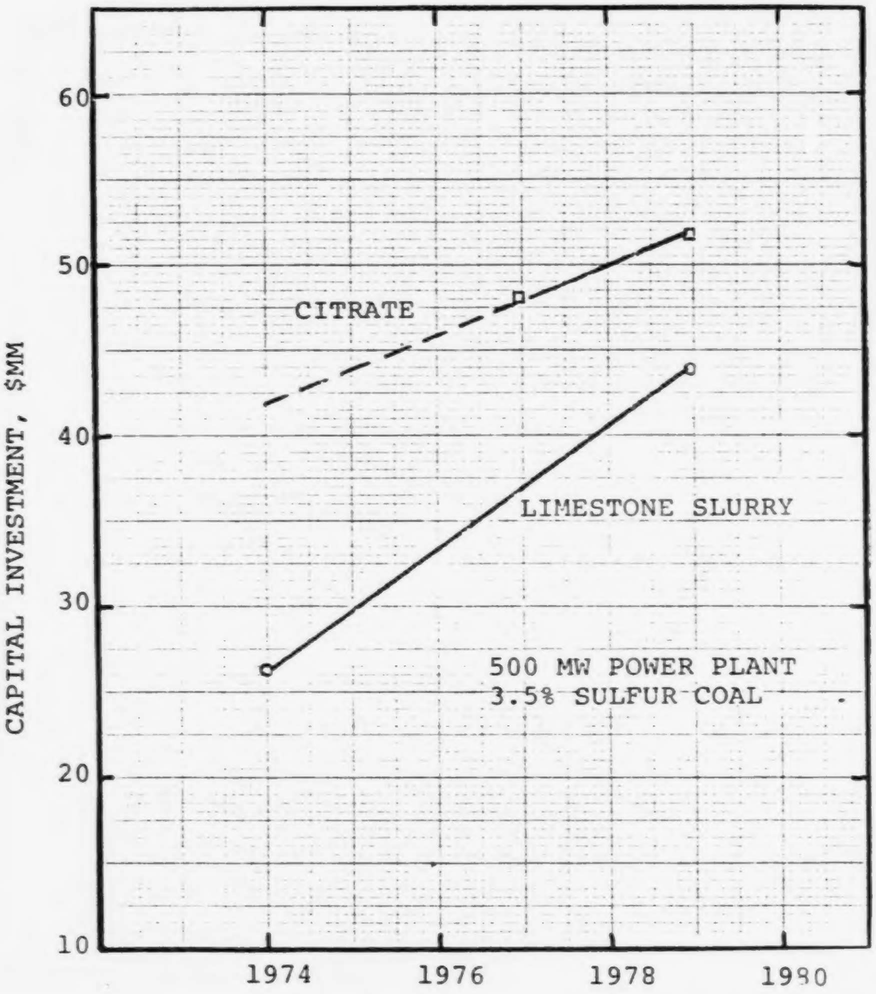

Figure 9-66. Capital Cost Comparison: limestone slurry vs. citrate process without H_2S generation.

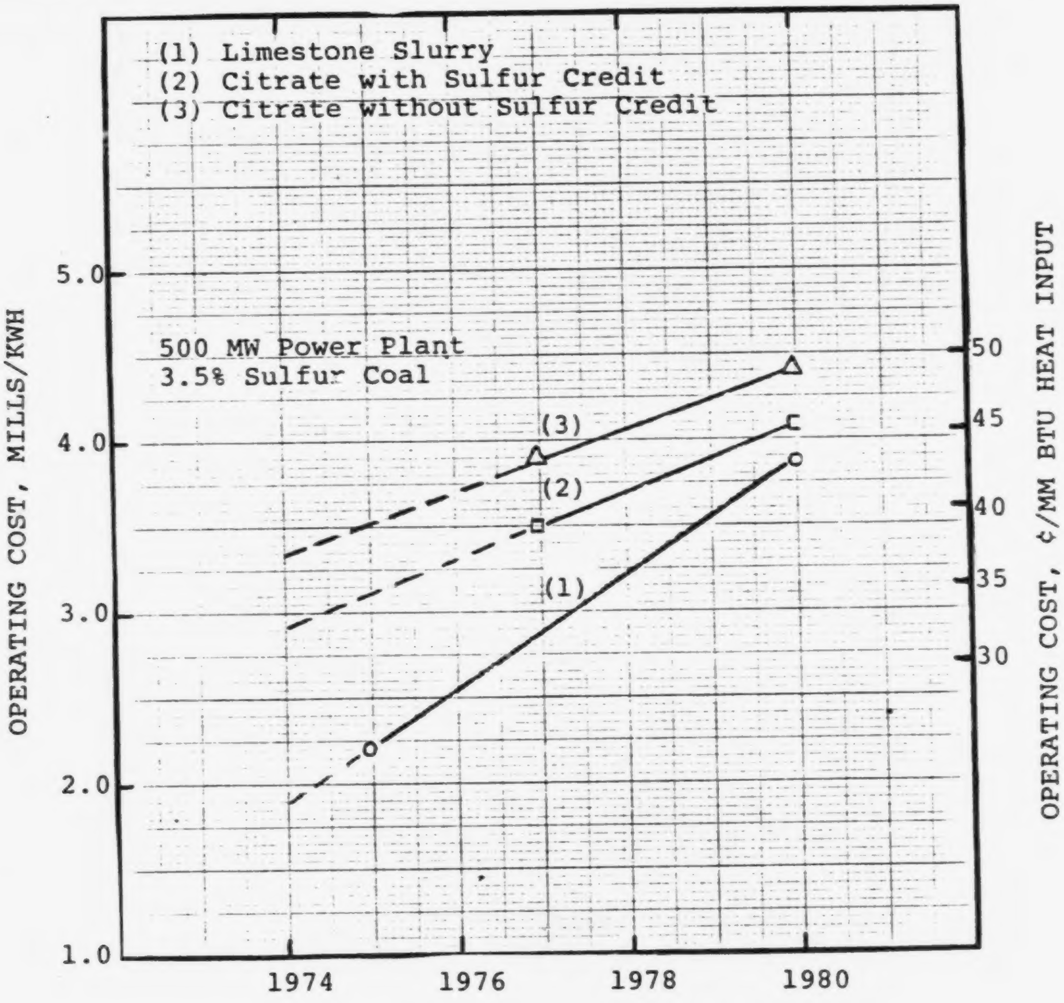

Figure 9-67. Operating Cost Comparison: limestone slurry vs. citrate process without H_2S generation.

Sulfur Dioxide Control for Coal Fired Boilers--
Sulfur dioxide may be controlled by FGD alone or by a combination of pyritic sulfur removal and FGD. To develop the operating costs for these systems the following basis was chosen:

Plant capacity	500 MW
Plant Heat rate	9,000 Btu/hr/KW
Heat input	4,500 MM Btu
Plant operation	90% Onstream (7,890 hrs/yr)
Cost basis	1977
Coal cost	$20/ton
Coal heat content	12,574 Btu/lb
Coal feed rate	357,882 lbs/hr
Sulfur in coal	3.02% (10,808 lbs/hr)

If the utility boiler at a coal conversion facility is fired with coal only, the following cases may be considered:

- Case 1: Sulfur dioxide control by flue gas desulfurization (FGD) alone.
- Cases 2 and 3: Remove 80 percent of the pyritic sulfur from the feed coal by heavy media washing and then apply FGD as necessary to meet sulfur dioxide emission standards.

Two coals were considered for this study. Their sulfur contents are shown as follows:

	Case 2 U.S. Average Coal	Case 3 Appalachian Coal
Pyritic Sulfur	1.91%	2.42%
Organic Sulfur	1.11	0.60
Total Sulfur	3.02%	3.02%

The Northern Appalachian coals are found in Maryland, Ohio, Pennsylvania and West Virginia.

The coal sulfur analysis used is typical for the middle and lower Kittanning coal bed. The organic sulfur content of these coals is lower than any other in the United States, with the exception of western coals which average about 0.45% organic sulfur.

When two coals have the same total sulfur content but the type of sulfur present varies, the coal with the higher pyritic sulfur content will be more amenable to a cleaning process such as heavy media washing.

The most stringent sulfur dioxide emission standard for coal fired boilers with a heat input greater than 250 MM Btu/hr is that of Wyoming at 0.2 lb/MM Btu. The next most stringent is that of New Mexico at 0.34 lb/MM Btu. The federal standard, and that of most states, is 1.2 lbs/MM Btu.

Figure 9-68 shows the operating costs for the three cases in ¢/MM Btu and $MM/yr vs. the SO_2 emission standard. The processing costs used are shown below (based on 1977):

- Heavy media washing: 29.1¢/MM Btu at 80 percent pyritic sulfur removal efficiency

- FGD by the citrate process without H_2S generating facilities: 40¢/MM Btu at 95.8 percent sulfur dioxide removal efficiency

The figure shows that FGD alone (Case 1) has a considerable advantage over coal cleaning plus FGD, particularly when meeting the more stringent emission standards. In the unlikely

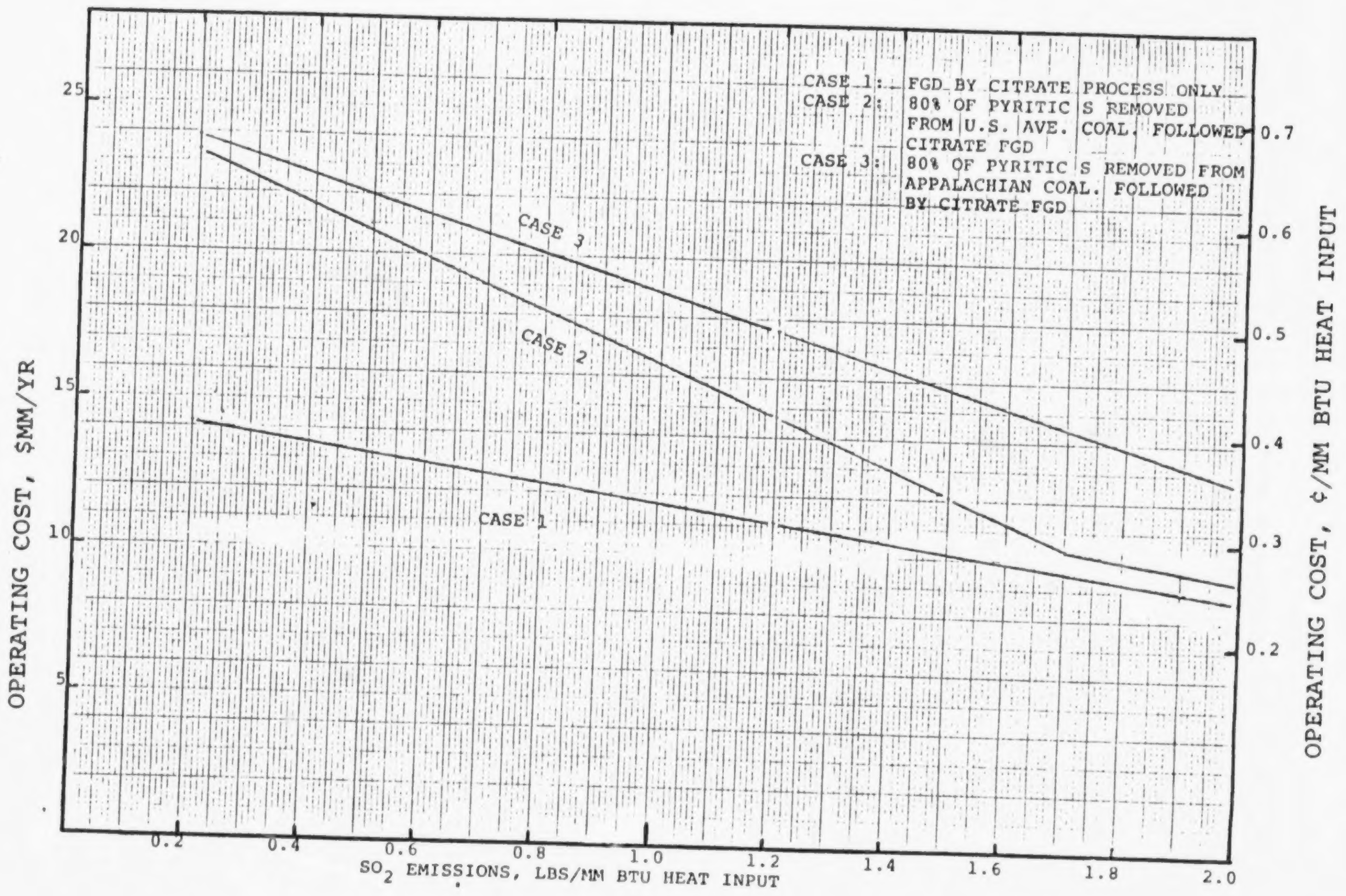

Figure 9-68. Sulfur dioxide control costs for coal fired boilers.

circumstance that the emission standards are relaxed, the combined cases, particularly Case 3 for Northern Appalachian coal, show less of a disadvantage.

Without FGD the sulfur dioxide emission from Case 2 would be 2.37 lbs/MM Btu and from Case 3, 1.72 lbs/MM Btu. The latter point is the reason for the break in the Case 3 line in Figure 9-68. Heating values of both coals are upgraded 6 percent by cleaning, from 12, 574 to 13,331 Btu/lb.

The following table shows the components of the treating costs, and the percentages of each treatment employed, to meet varying SO_2 emission standards.

	Operating Cost, ¢/MM Btu		
SO_2, Lb/MM Btu	0.2	1.2	2.0
Case 1 (Scrubbing)	40.0(100.0%)	31.3 (78.3%)	24.4(60.9%)
Case 2			
Washing	29.1(100.0%)	29.1(100.0%)	29.1(100.0%)
Scrubbing	38.2(95.6%)	20.6(51.6%)	6.5(16.4%)
Total	67.3	49.7	35.6
Case 3			
Washing	29.1(100.0%)	29.1(100.0%)	26.3(90.3%)
Scrubbing	36.8(92.1%)	12.3(30.7%)	0.0(0.0%)
Total	65.9	41.4	26.3

Therefore, the conclusion may be drawn that FGD alone appears to be the most economical method to meet stringent sulfur dioxide emission standards of 0.2 to 1.2 lb/MM Btu for coal fired boilers.

Sulfur Dioxide Control for Oil Fired Boilers--
Operating costs were developed for control of sulfur dioxide emissions from oil fired boilers in the same manner as for coal fired boilers, by considering FGD alone and also in combination

with hydrodesulfurization of the oil. The basis for the study was:

Plant Capacity	50 MW
Plant heat rate	9,000 Btu/hr/KW
Heat input	4,500 MM Btu/hr
Plant operation	90% Onstream (7,890 hrs/yr)
Cost basis	1977
Oil type	Residual
Oil heat content	6.24 MM Btu/barrel
Oil feed rate	17,310 Barrels per day
Sulfur in oil feed	1.81% (60% of sulfur in coal feed)

For Case 1, hydrodesulfurization costs for 92 percent sulfur removal, with an allowance of 25 percent of the cost as credit for salable sulfur, were calculated to be $1.82 per barrel of product oil (equivalent to $1.73 per barrel of feed oil) or $0.292/MM Btu of heat input to the boiler. Sulfur in the residual oil feed was 4441 lbs/hr. If all the oil were desulfurized, the the desulfurized oil to the boiler would contain 355 lb/hr of sulfur which, when burned would yield a flue gas containing 0.16 lb/MM Btu of sulfur dioxide. For the various flue gas compositions studied, varying amounts of residual oil were desulfurized. The resulting costs are shown in Figure 9-69.

For Case 2, costs for the citrate FGD process were developed in the same manner as for the preceding study on sulfur dioxide control for coal fired boilers. Corresponding costs are shown in Figure 9-69.

Inspection of Figure 9-69 leads to the conclusion that for oil fired boilers hydrodesulfurization of the feedstock is a more economical method for control of sulfur dioxide emissions than is FGD and, as the maximum allowable sulfur dioxide content of the flue gas is reduced, the advantage increases.

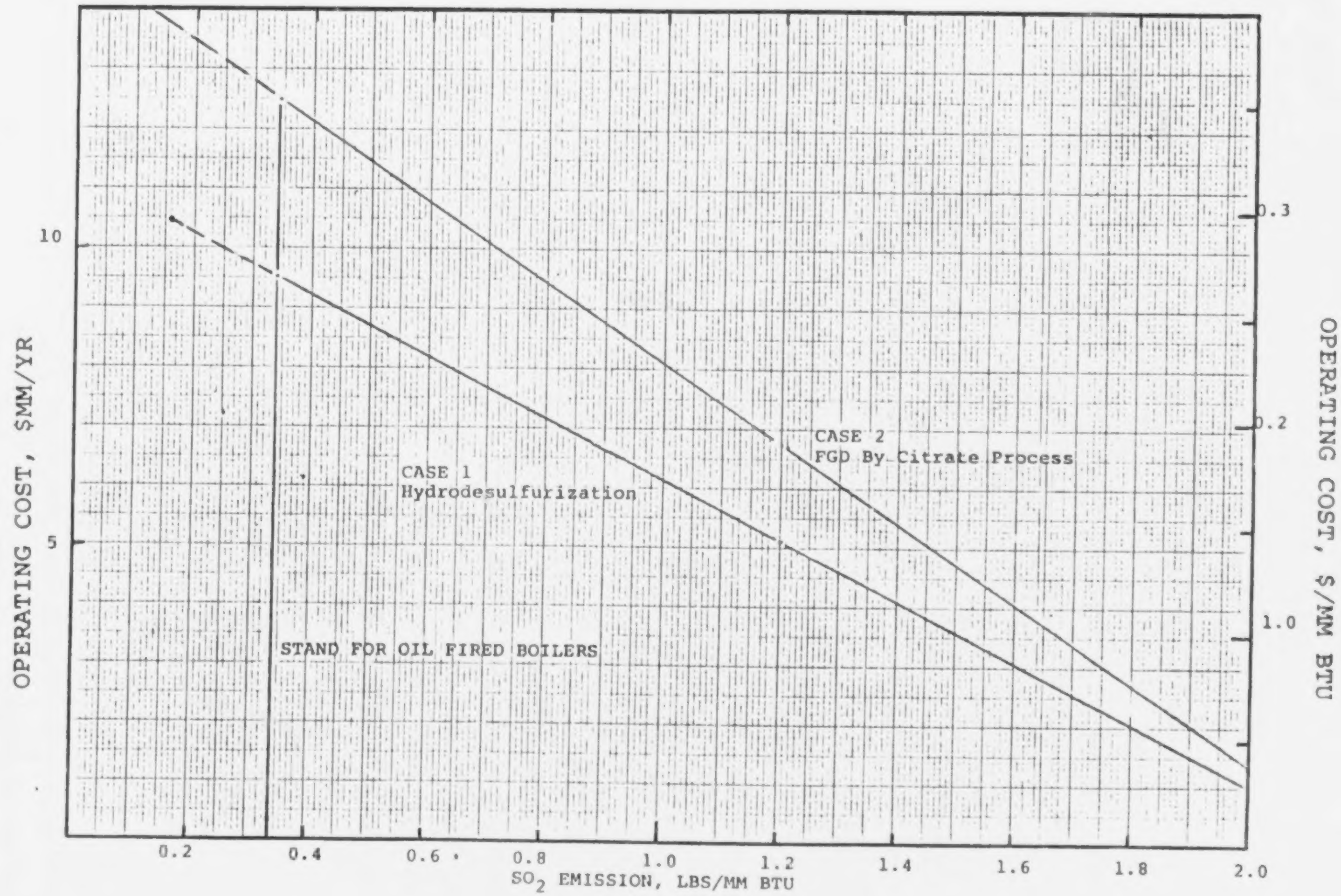

Figure 9-69. Sulfur dioxide control costs for oil fired boilers.

Nitrogen Oxides Control

NO_x Control by Boiler Modifications--

It is generally recognized that the first step taken in the direction of lowering NO_x emission from fossil fuel fired boilers will be boiler modifications. Design changes which lower the flame temperature and reduce oxygen availability result in lower NO_x formation. These changes can be one or more of the following:

- o Two-stage combustion
- o Low excess air firing
- o Flue gas recirculation

The approximate percentage reduction of NO_x emissions by the different techniques is shown below.

	Percent NO_x Reduction		
	Coal	Oil	Gas
Staged Combustion	35	35	45
Low Excess Air	20	25	25
Combined Staged Combustion and Low Excess Air	40	40	50
Flue Gas Recirculation	25	25	45

Thus, it is concluded that a 40 to 50 percent reduction in NO_x emissions can be achieved by boiler modifications.

The most stringent standards for NO_x emissions from fuel combustion at present are those of New Mexico. These standards along with projected future standards for NO_x emissions are:

	New Mexico NO_x Standard		Projected Future NO_x Standards	
			1980	1985
	lb/MM Btu	ppm(v)	ppm(v)	ppm(v)
Coal	0.45	338	200	100
Oil	0.30	225	150	90
Gas	0.20	150	150	50

The table which follows gives expected NO_x emissions for both uncontrolled and controlled boiler conditions:

	Expected NO_x Emissions		
	Coal	Oil	Gas
Uncontrolled, ppm	500	280	200
Controlled, ppm	370	150-210	85-110
NO_x Reduction with Controls, %	26	25-46	45-57

It may be concluded that boiler modifications are sufficient to meet the most stringent present standards for NO_x, perhaps with coal being borderline. These modifications also appear to meet the projected 1980 goals for oil and gas, but not for coal. However, none of the projected 1985 goals can be met using only boiler modifications.

Boiler modifications are believed to be the lowest cost NO_x control measure. However, neither investment costs nor operating costs for these changes were available. Therefore, they are not incorporated in the following sections which address economics of the control of NO_x. For example, in these sections, a coal fired boiler is assumed to have 500 ppm(v) NO_x in the flue gas and an oil fired boiler is assumed to have 280 ppm(v).

NO_x Control by Flue Gas Denitrification--

About 40 to 50 flue gas denitrification and denitrification/-desulfurization processes are reported in the literature. They are both the wet type, which normally removes both NO_x and SO_2, and the dry type. A report by TVA studied these processes and recommended eight for further study (11, p.370). Included in these are the following processes for which some preliminary economic data was available:

- o UOP - Shell (SO_2/NO_x)
- o UOP - Shell (NO_x only)
- o Asahi Chemical (SO_2/NO_x)

Figure 9-70 shows estimated capital investment costs of these processes as a function of capacity in MW. An exponent of 0.7 was assumed to prepare the figure. It should be noted that the flue gas flow is assumed to be 330 moles per hour (MPH) per MW. Therefore, a 500 MW power plant will produce about 165,000 MPH of flue gas.

Capital investment costs for the three processes (1977) for operation in a 500 MW power plant are reported to be (11):

	$/KW
UOP-Shell (SO_2/NO_x)	131
Aashi Chemical (SO_2/NO_x)	127
UOP Shell (NO_x)	31

Operating costs for the processes are shown below:

	Operating Cost*	
	Mills/KWH	$/MM Btu
UOP-Shell (SO_2/NO_x)	5.0	0.5555
Asahi Chemical (SO_2/NO_x)	7.4	0.8222
UOP-Shell (NO_x)	1.4	0.1555

*Heat rate assumed to be 9,000 Btu/hr/KW

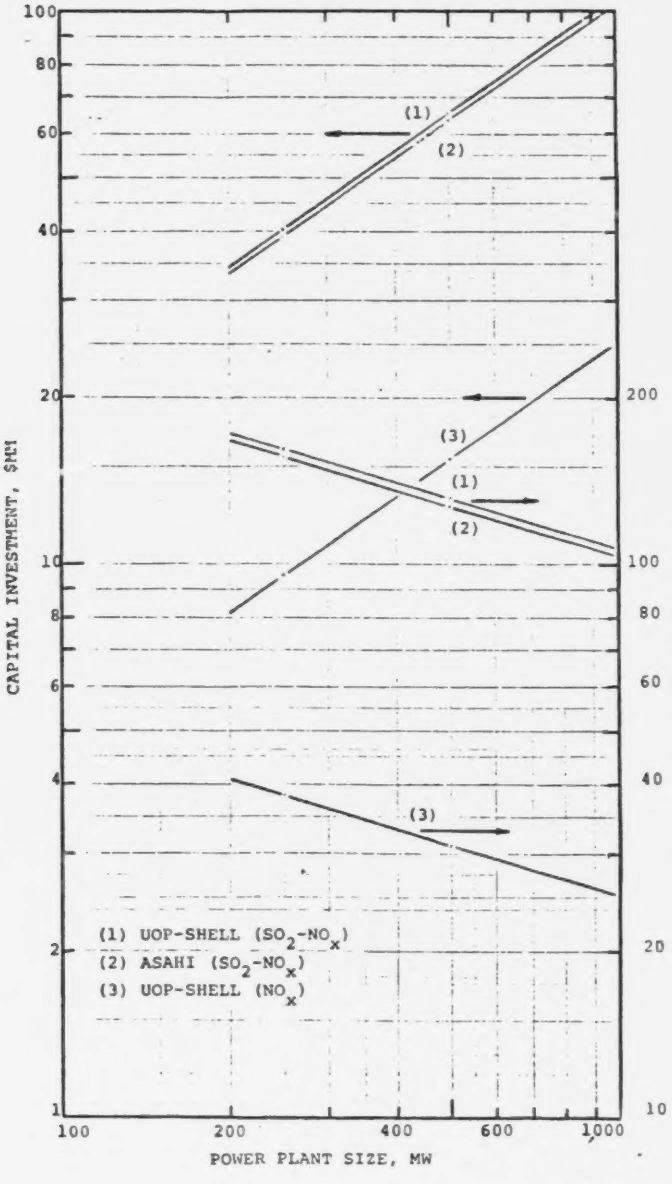

Figure 9-70. Capital investment for NO_x/SO_2 control.

These costs are based on treating a flue gas containing 2,580 ppm(v) SO_2 and 634 ppm(v) NO_x for the UOP-Shell processes and 2200 ppm(v) SO_2 and 600 ppm (v) NO_x for the Asahi process.

Since the UOP-Shell process (SO_2/NO_x) has a considerably lower reported operating cost than the Asahi process, it was selected for further study. For NO_x removal only, the UOP-Shell process was also selected.

NO_x Control for Oil Fired Boilers--
If a residual fuel such as tar, tar oil, and oil skimmings produced at a coal conversion plant and containing, typically about 0.4 percent N, is burned in a utility boiler, the uncontrolled NO_x emissions are expected to be:

	ppm(v)
Fuel NO_x	207 (45% conversion)
Thermal NO_x	73
Total	280

These emissions exceed even the present most stringent standards, not to mention projected future standards. Therefore, some type of control will be required. Three alternate strategies were examined:

- o Case 1: Flue gas denitrification alone
- o Case 2: Fuel hydrotreating alone
- o Case 3: A combination of 100 percent fuel hydrotreating plus the necessary degree of flue gas denitrification to meet a given standard

Costs of the treatment methods are:

- o FG Denitrification = \$0.1555/MM Btu at 90 percent NO_x removal (based on UOP-Shell process)
- o Fuel Hydrotreating = \$0.2918/MM Btu at 90 percent nitrogen removal efficiency

The operating costs for each of these cases are plotted as a function of the NO_x emission for each of the cases in Figure 9-71. The basis for these costs was:

Cost basis		End 1977
Plant capacity		500 MW
Plant heat rate		4,500 MM Btu/hr
On-stream factor		90%
UOP-Shell NO_x removal:	efficiency	90%
	cost	$0.1555/MM Btu
Hydrodenitrification:	efficiency	90%
	cost	$0.2918/MM Btu

After examination of the information available the following conclusions may be drawn:

- Case 1, flue gas denitrification alone, is far superior to the other cases. The proposed 1980 standard of 150 ppm NO_x can be met by treating 51.6 percent of the flue gas at a cost of 8.02¢/MM Btu. The proposed 1985 standard of 90 ppm NO_x can be met by treating 75.4 percent of the flue gas at a cost of 11.7¢/MM Btu. This system has an ultimate limit of 28 ppm NO_x with full flue gas treatment at a cost of 15.55¢/MM Btu.
- Case 2, fuel hydrotreating alone, is costly and limited in the amount of NO_x reduction attainable. If the original fuel contains 0.4 percent N and 80 percent reduction is realized, or 0.8 percent N and 90 percent reduction, to a fuel N content of 0.08 percent, the following NO_x emission is expected:

	ppm(v)
Fuel NO_x	65 (70% conversion)
Thermal NO_x	73
Total	138

This method is handicapped to a certain extent since

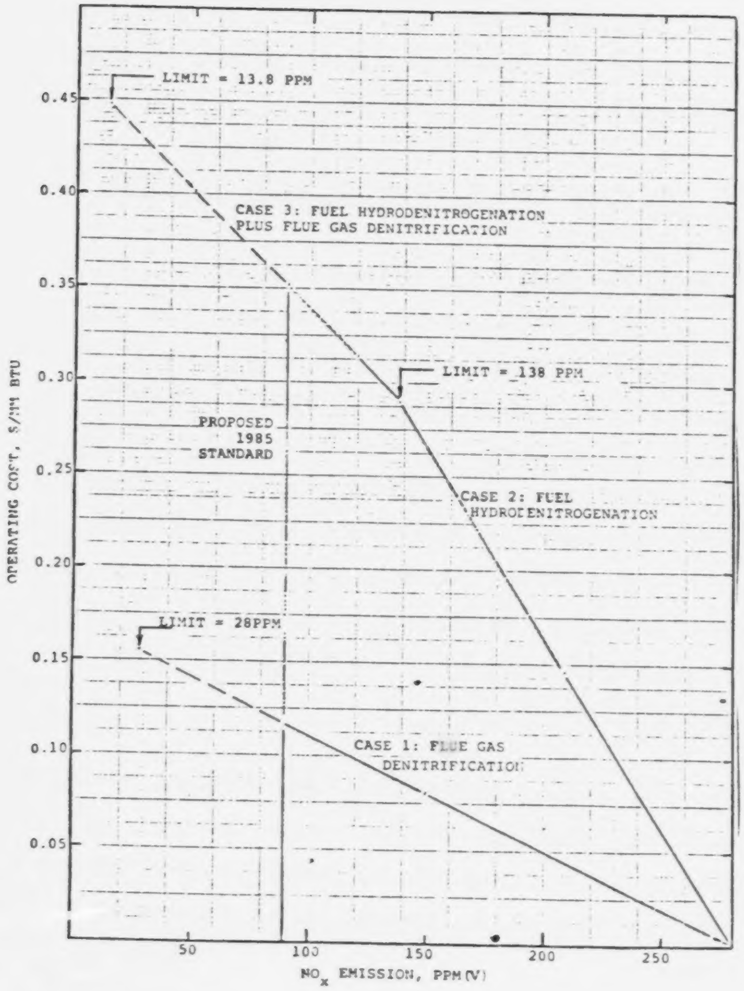

Figure 9-71. Operating costs for NO_x control for oil fired boilers.

fuels with a lower nitrogen content exhibit a higher conversion to NO_x.

- Case 3, the combination of 100 percent fuel hydrotreating and partial to complete flue gas denitrification, is the most costly of the methods examined. A lower ultimate NO_x level can be achieved, however. This limit is expected to be about 14 ppm at a cost of 44.7¢/MM Btu.

Combined Sulfur Dioxide and Nitrogen Oxide Control

Control of SO_2/NO_x by Fuel Oil Hydrotreating--

Hydrotreating of gas oils (boiling range 650 to 1,100°F) and residual oils (boiling range >650°F) is a widely practiced scheme for desulfurization of these fuels. Sulfur present is converted to hydrogen sulfide which is then fed to a Claus unit for elemental sulfur production. Sulfur removal efficiencies of 97 percent for gas oils and 92 percent for residual oils are reported (6, p.5). Hydrodenitrogenation occurs simultaneously as nitrogen in the feed is converted to ammonia. A nitrogen removal efficiency of about 80 percent, perhaps as high as 90 percent, can be attained (7, pp.15,17).

In Figure 9-72 are shown the estimated capital investments (1977) and in Figure 9-73, the operating costs for gas oil and residual oil hydrotreating as a function of feed capacity. The basis for development of the operating costs is as follows:

Cost Basis	End 1977
Plant capacity	500 MW
Plant heat rate	4,500 MM Btu/hr
On-stream factor	90%
Product/feed	0.947

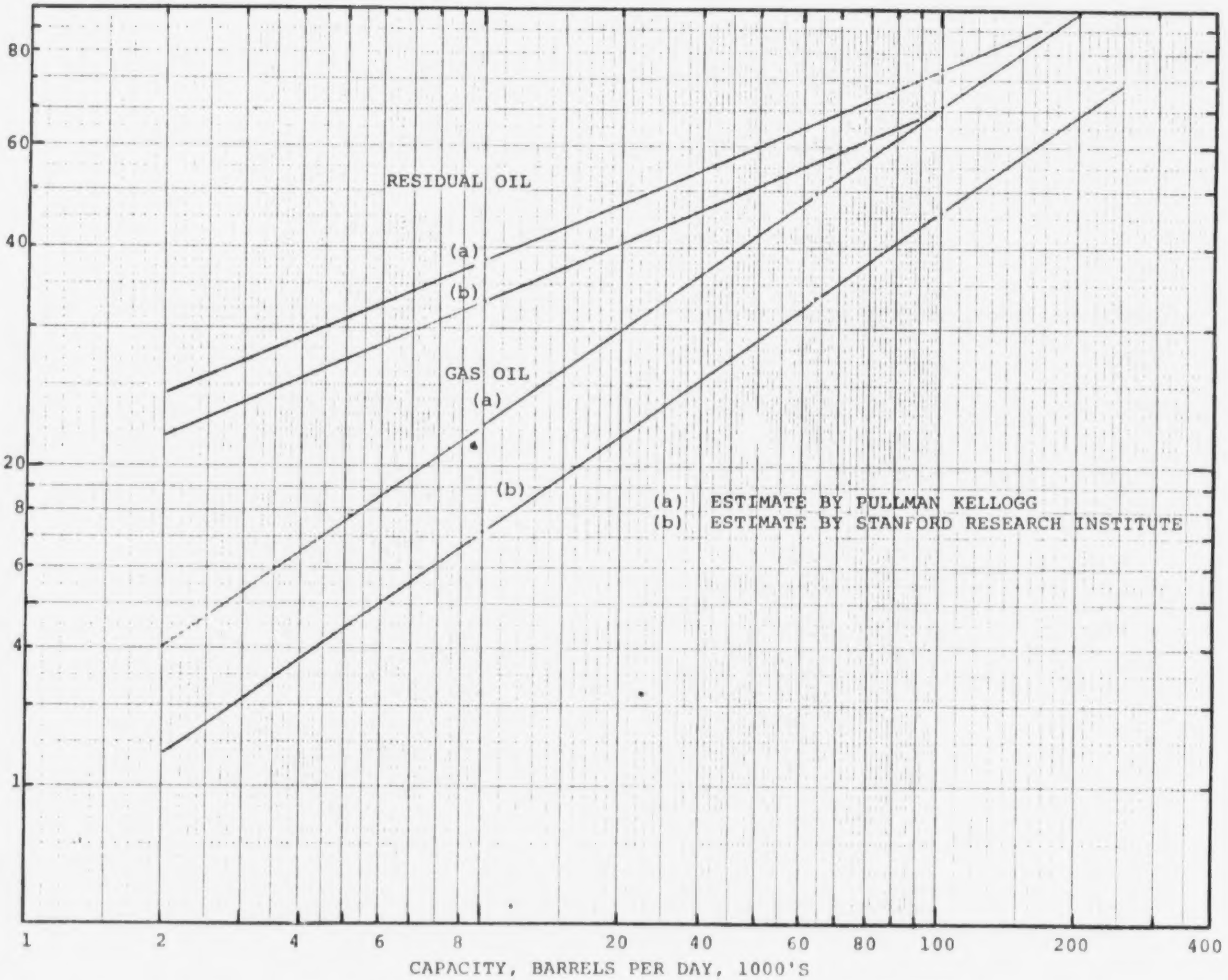

Figure 9-72. Capital investment for fuel oil hydrotreating.

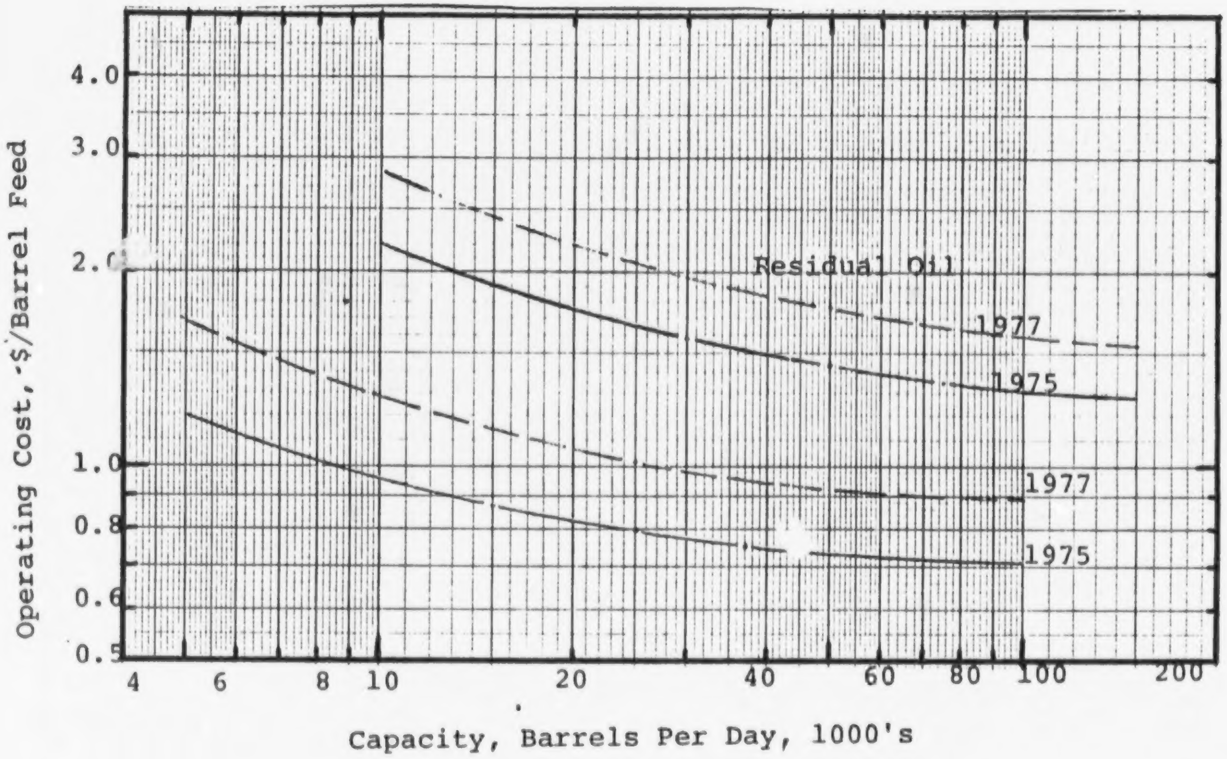

Figure 9-73. Operating costs for fuel oil hydrotreating.

No credit was allowed in development of the costs for byproduct naphtha and fuel gas. If the credit were allowed, operating costs would decrease about 25 percent.

For each capacity a range of costs is shown in Figure 9-72, of which the upper line is derived from Pullman Kellogg estimates and the lower from Stanford Research Institute estimates. Sulfur in feed and product to hydrotreating are:

	Feed	Product	Removal
Residual oil	3.90%	0.3%	92%
Gas oil	3.25	0.1	97

For calculation of capital costs at various capacities an exponent of 0.6 was used for residual oil and 0.7 for gas oil.

A large coal conversion facility of 250 billion Btu/day SNG requires a utility boiler which is about equivalent to a 500 MW power plant. If the plant is assumed to have a heat rate of 9,000 Btu/hr/KW, then 4,500 MM Btu/hr of fuel are needed. Tars and tar oils produced in low temperature gasification processes have a heating value of about 16,500 Btu/lb. The specific gravity of these materials is about 1.08 resulting in a heating value of about 148,600 Btu/gallon or 6.24 MM Btu/barrel (42 gallons/-barrel). To supply 4,500 MM Btu hr, 721.1 barrels/hr or 17,310 barrels/day of fuel are needed. The capital investment cost (1977) of a residual oil hydrotreater of this capacity is estimated to be in the range of $29 to $35 MM. The operating cost is expected to be about $2.30/barrel feed without credit for byproducts. After taking a 25 percent credit for byproducts and adjusting for the slightly reduced output of 94.7 percent of the feed, the operating cost becomes $1.82/barrel of product or $0.2918/MM Btu.

Control of SO_2 and NO_x for Oil Fired Boilers--

In previous sections, SO_2 and NO_x control for oil fired boilers were addressed separately. In this section, the combined control of these pollutants will be discussed. Three cases were studied:

- Case 1: Hydrotreat only. With full hydrotreatment the ultimate SO_2 emission is 0.176 lb/MM Btu and the ultimate NO_x emission is 138 ppm.
- Case 2(a): Hydrotreat 91.9 percent of the fuel oil sufficient to meet an SO_2 emission standard of 0.34 lb/MM Btu. Remove NO_x to varying levels with UOP-Shell process.
- Case 2(b): Hydrotreat 97.8 percent of the fuel oil sufficient to meet an SO_2 emission standard of 0.22 lb/MM Btu. Remove NO_x to varying levels with the UOP-Shell process.
- Case 3: No hydrotreating. Combined SO_2/NO_x removal by UOP-Shell process.

Figure 9-74 is a plot of operating cost vs. SO_2 and NO_x emissions for the cases. TABLE 9-30 summarizes emissions and costs for the cases.

Costs were developed from the following basis:

Cost basis		End 1977
Plant capacity		500 MW
Plant heat rate		4500 MM Btu/hr
On-stream factor		90%
Fuel hydrotreating:	sulfur removal	92%
	nitrogen removal	90%
	cost	\$0.2918/MM Btu
UOP-Shell NO_x removal		90%
	cost	\$0.1555/MM Btu
UOP-Shell SO_2/NO_x removal:		
	SO_2/NO_x removal	90%
	cost	\$0.5555/MM Btu

TABLE 9-30. CONTROL OF NO_x/SO_2 FOR OIL FIRED BOILERS

	Case 1 Hydrotreat Only	Case 2(a) Hydrotreat 91.9% plus UOP-Shell NO_x Process			Case 2(b) Hydrotreat 97.8% plus UOP-Shell NO_x Process			Case 3 UOP-Shell SO_2/NO_x Process
Full Hydrotreat (100%)								
SO_2, lb/MM Btu	0.176							
NO_x, ppm	138							
Cost, $/MM Btu	0.292							
Partial Hydrotreat (97.8%)								
SO_2, lb/MM Btu	0.220				0.220	0.220	0.220	
NO_x, ppm	141				90	43	28	
Cost, $/MM Btu	0.285				0.348	0.404	0.424	
Partial Hydrotreat (91.9%)								
SO_2, lb/MM Btu	0.340	0.340	0.340	0.340				
NO_x, ppm	148	90	43	28				
Cost, $/MM Btu	0.268	0.336	0.391	0.409				
Full Flue Gas Treat (100%)								
SO_2, lb/MM Btu								0.220
NO_x, ppm								28
Cost, $/MM Btu								0.555
Partial Flue Gas Treat (93.9%)								
SO_2, lb/MM Btu								0.340
NO_x, ppm								43
Cost, $/MM Btu								0.522

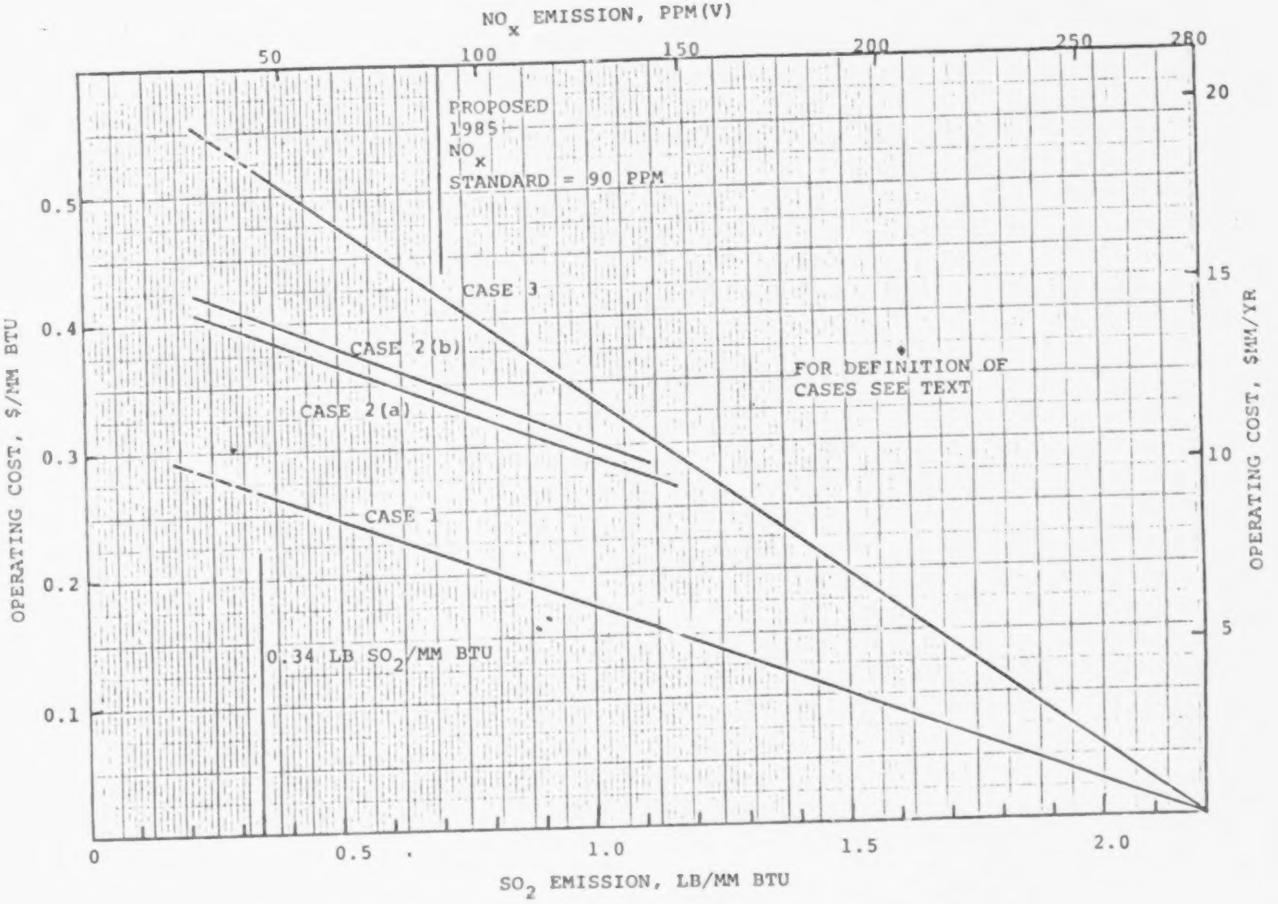

Figure 9-74. Operating costs for NO_x/SO_2 control for oil fired boilers.

The following conclusions may be drawn:

- To meet an SO_2 emission standard of 0.34 lb/MM Btu and the proposed 1980 NO_x goal of 150 ppm, Case 1 with 91.9 percent hydrotreating is clearly superior, with a cost of $0.268/MM Btu
- To meet the same SO_2 standard and the proposed 1985 NO_x goal of 90 ppm, Case 2(a) is preferred with 91.9 percent hydrotreating and UOP-Shell NO_x and a cost of $0.336/MM Btu. It is possible that a combination of hydrotreating and boiler modifications (which reduce NO_x emissions by 30 to 40 percent) may suffice to meet the 1985 goal. If so, this case would cost less than Case 2(a). The NO_x emission can be lowered to 43 ppm by treating more flue gas, but with a cost of $0.391/MM Btu.
- Case 3 does not appear to be competive with Case 2(a). To meet SO_2 standards of 0.34 lb/MM Btu and 43 ppm NO_x costs $0.522/MM Btu.

Control of SO_2/NO_x for Coal Fired Boilers--

Two cases were analyzed for control of SO_2 and NO_x from coal fired boilers:

- Case 1: Removal of NO_x by the UOP-Shell process and removal of SO_2 by the citrate process
- Case 2: Removal of both SO_2 and NO_x by the UOP-Shell process

Costs were developed for the two cases on the following basis:

Cost Basis	End 1977
Plant capacity	500 MW
Plant heat rate	4500 MM Btu/hr
On-stream factor	90%
SO_2/NO_x removal efficiency	90%

Operating costs:	UOP-Shell NO_x	\$0.1555/MM Btu
	Citrate SO_2	\$0.3889/MM Btu
	UOP-Shell SO_2/NO_x	\$0.5555/MM Btu

If all of the flue gas is treated for both cases with 90 percent removal efficiency for both SO_2 and NO_x, the costs and emissions for the two cases are:

	Case 1	Case 2
Operating cost, \$/MM Btu	0.5444	0.5555
SO_2, lb/MM Btu	0.48	0.48
NO_x, ppm(v)	50	50

As shown in Figure 9-75, no clean-cut choice can be made based on operating costs. The SO_2 emissions obtained are adequate for most states, but they exceed the most stringent standard of 0.2 lb/MM Btu set by Wyoming. The NO_x emissions are lower than the projected 1985 goal of 100 ppm(v).

If 88.9 percent of the flue gas is treated, so that the NO_x emissions are 100 ppm for both cases, the operating costs and emissions are:

	Case 1	Case 2
Operating cost, \$MM/Btu	0.484	0.494
SO_2, lb/MM Btu	0.96	0.96
NO_x, ppm(v)	100	100

Again, the operating costs are too close to choose between the cases.

The SO_2 emissions meet the present standards set by most states, but obviously not the most stringent standards.

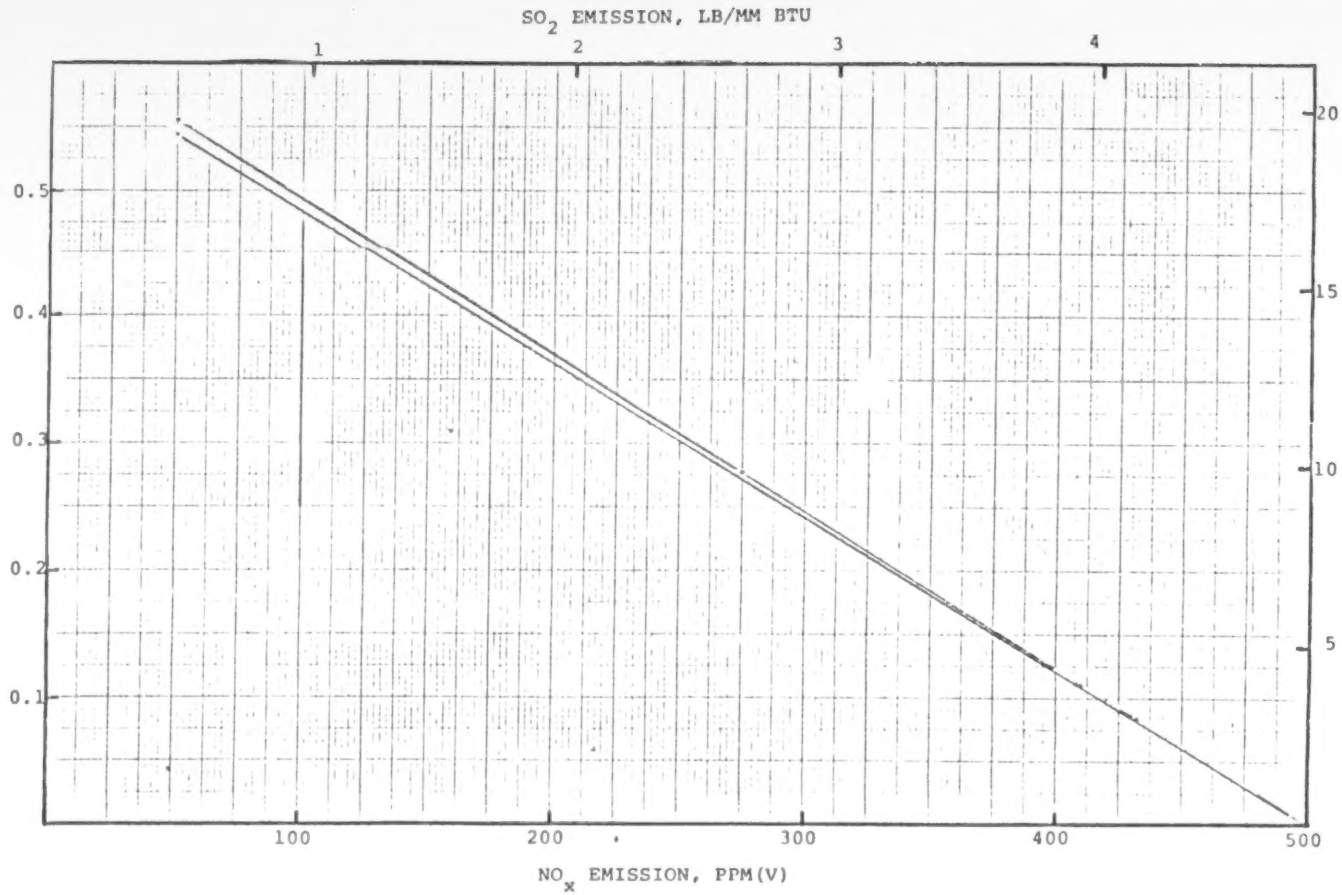

Figure 9-75. Operating costs for SO_2/NO_x control for coal fired boilers.

In order to meet the most stringent standards of 0.2 lb SO_2/MM Btu and 100 ppm NO_x, Case 1 appears to be required because of its higher SO_2 removal capability. The costs and emissions to meet these standards are obtained by partial treatment of the flue gas, before the air heater, by the UOP-Shell process and full treatment of the flue gas after the ESP by the citrate process operated to obtain 95.8 percent removal efficiency:

	Case 1
Operating cost, $/MM Btu	
UOP-Shell NO_x	0.1382
Citrate	.40
Total	0.5382
SO_2: lb/MM Btu	0.2
NO_x: ppm(v)	100

If Case 2 can be operated to achieve 95.8 percent SO_2 removal, it will probably serve as well.

Production of Elemental Sulfur from H_2S and Control of Organic Sulfur Emissions

The Claus process and the Stretford process are most nearly universally applicable in converting H_2S-rich gas from a solvent system in a coal conversion plant to elemental sulfur. Neither of these processes produces a tail gas that is environmentally acceptable, a circumstance that has led to combination processes that minimize pollution from H_2S, COS, CS_2, and SO_2:

Processes Combined with Claus	Processes Combined with Stretford
Beavon	Hot Carbonate
SCOT	Holmes-Maxted
ARCO	Carpenter-Evans
Incineration	British Gas Council
Lucas	Iron Oxide

Selection of the optimum processing technique can depend on a number of factors, including:

- The relative amounts of CO_2 and H_2S in the acid gas (CO_2/H_2S ratio)
- The presence of organic sulfur compounds (COS, CS_2) in the feed gas
- The presence of ammonia and hydrocarbons in the feed
- The volumetric flow of acid gas
- The concentration of H_2S
- The emission standards in effect for various sulfur species
- Reducing gas cost and availability
- Relative costs of raw materials, utilities and manpower
- Capital related costs

It was beyond the scope of this task to evaluate all of these variables. Some economic data for the Claus/Beavon combination were supplied by the Ralph M. Parsons Co. Capital costs for these plants are plotted as a function of capacity in Figure 9-76, using the exponent 0.6.

The H_2S concentration in the feed is shown to have a considerable effect on the investment required. Decreasing the H_2S content from 40 to 10 mol percent results in a 70 percent increase in capital investment for the same sulfur production. Therefore, it is clear that a concentrated H_2S feed is highly desirable.

Although no specific economic data were supplied for the other processing schemes, Parsons stated that a Claus/ARCO unit typically costs about 15 percent more than Claus-Beavon.

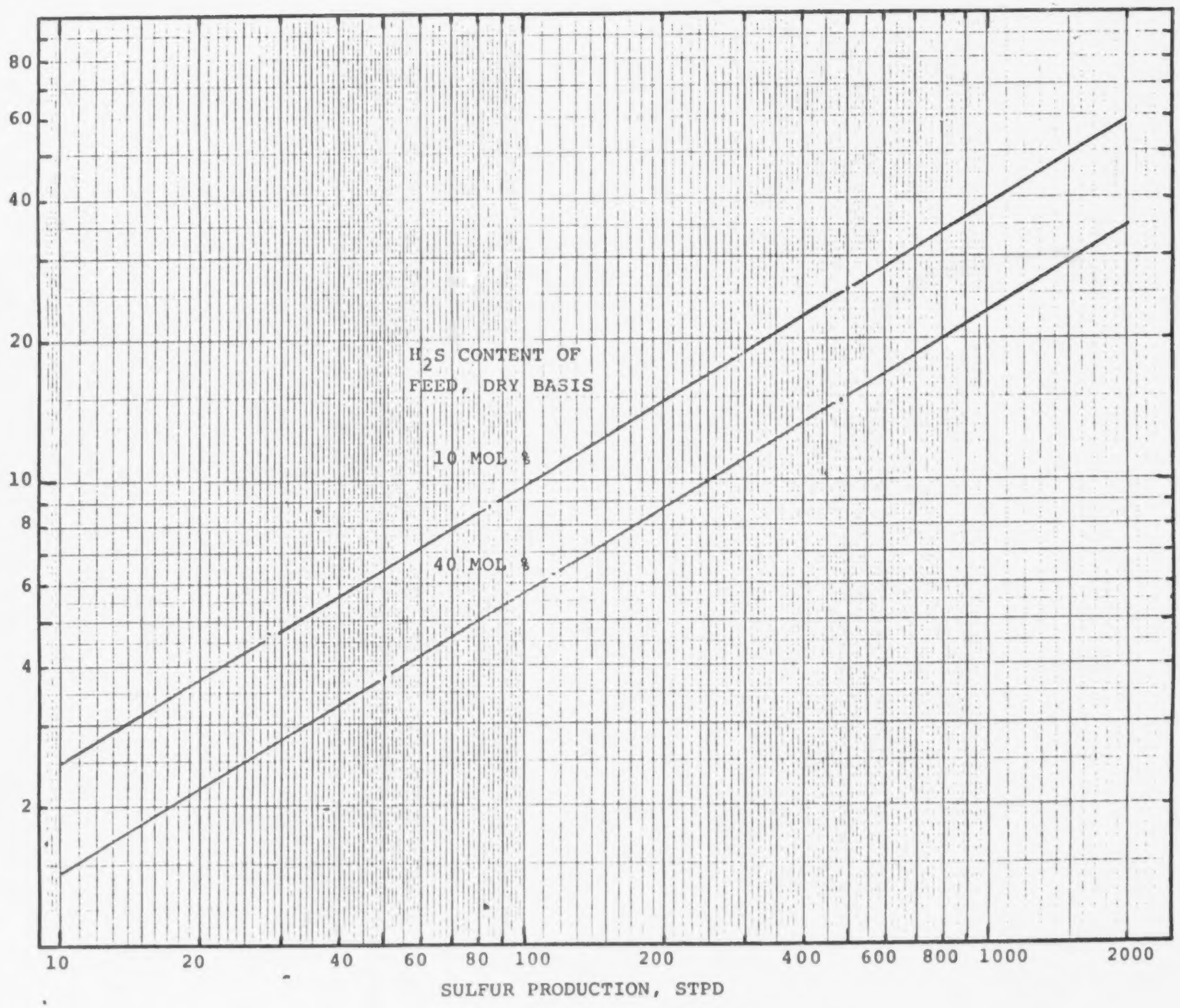

Figure 9-76. Capital investment for Claus-Beavon sulfur recovery.

Estimated operating costs for the Claus-Beavon combination are shown in Figure 9-77, developed with the following criteria:

Cost basis	1977
On-stream factor	90%
Capital charges	25%
Steam credit	$3/1000 lb
Fuel cost	$3/MM Btu
Power cost	$0.025/KWH
Labor cost	2 men/shift ($480/day)

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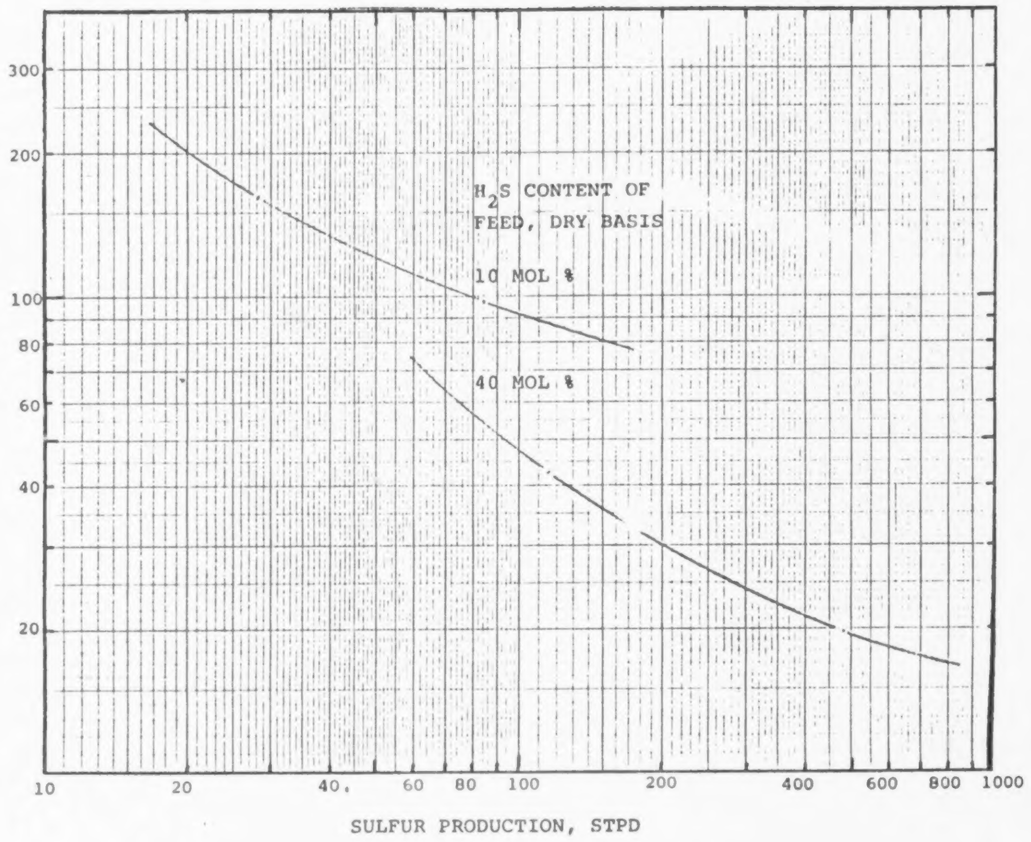

Figure 9-77. Operating costs for Claus-Beavon sulfur recovery.

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NEED FOR ADDITIONAL DATA, INFORMATION AND DEVELOPMENT

Coal Pretreatment

Sulfur must be removed from coal before conversion or be recovered in the conversion processes. Physical pyritic sulfur removal by, for example, heavy media washing, is incomplete but relatively inexpensive, and the sulfur values are usually thrown away. Chemical pyritic sulfur removal by, for example, the Meyers process, is effective in reducing sulfur and yields salable sulfur and iron sulfate that would probably be thrown away. Neither process is effective in removing organic sulfur. Use of either type would reduce the requirement for sulfur recovery and sulfur dioxide control in the coal conversion processes.

Development of economic data on coal cleaning methods and on the facilities within the coal conversion process is required for the making of meaningful decisions to:

- Reduce sulfur by physical coal cleaning and recover the rest in the conversion plant

- Reduce sulfur by physical coal cleaning, recover part in the conversion plant and throw away the rest as FGD sludge

- Reduce sulfur by chemical coal cleaning, recover or thow away the sulfur and recover or throw away the remaining coal sulfur in the conversion plant

More information is needed on operability of the Meyers process.

Particulates

Further study and evaluation are needed on the quantities and compositions, including particulates, of gases released from coal feeding devices such as lock hoppers.

More data are needed on particulate collection efficiencies and costs as applied to such coal dust control devices as cyclones, baghouses and electrostatic precipitators.

Data on quantity and particle size distribution of dusts evolved in coal crushing, sizing, pulverizing and drying steps would be helpful in developing control methods and meaningful economics.

Ash Quench

While much data and information are available on gases evolved from quenching in Lurgi Dry Ash gasification, little real data have been collected on the gases evolved from other processes. If, as in many conceptual designs, ash is quenched with contaminated water, the contaminants themselves may volatilize, together with products of reactions of the contaminants with the ash, or the ash with water.

Data are needed on quantity and composition of the ash quench streams from the various conversion processes and the variations for any one process as the quench water composition varies.

Particulate Removal from Gasifier Offgas

Particulates are carried in the gasifier offgas as ash, char, and unreacted coal. When the gas stream is water quenched for cooling and tar condensation purposs, as in the Lurgi process, the particulates are removed from the gas stream by the quench water. Since the Lurgi process has been in commercial operation for many years, the problems of particulate separation from the liquid streams have reached satisfactory solutions.

In the high temperature slagging gasifiers little or no phenols, oils and tars are produced and direct quenching of the raw gas stream for cooling and condensation is not practiced. In the Parsons conceptual design for a liquefaction plant, for example, there are two slagging gasifiers, one oxygen blown to produce process gas and the other air blown to produce fuel gas for process heating and steam generation. In both of these, the raw gas carries char and the char is mechanically separated from the gas streams by combinations of cyclones, dust filters and electrostatic precipitators. While it is permissible to stipulate particulate removal efficiency in a conceptual design for process purposes, final engineering design of a full scale conversion plant requires as exact a set of data as can be assembled.

Indications from the literature, concerning investigations on high temperature, high efficiency, high volume cyclone designs (which parameters are usually considered as being mutually exclusive), are that much more data are needed on performance at expected operating conditions. The same impressions are gained concerning hot gas filters. Electrostatic precipitator performance on hot fly ash is well documented and, given the dust loading, resistivity and operating conditions, design should offer no unusual difficulty.

Acid Gas Removal

While removal of hydrogen sulfide and carbon dioxide from the process gas stream is not usually considered to be an emissions problem, the composition of the hydrogen sulfide stream affects the performance of the sulfur recovery system, and thus affects the treatment that must be applied so that the final vent gas stream will meet environmental standards. In a like manner, the composition of the carbon dioxide stream affects the performance of downstream process steps and thus affects the vent gas treatment step.

Suggested avenues of investigation in several of the acid gas removal processes are:

- Selexol: Study effect of feed acid gas composition on removal efficiency at various operating temperatures and pressures
- Rectisol: Determine the solvent retention of heavy hydrocarbons and the composition and quantity of the fugitive carryover from the process at high pressures
- Monoethanolamine (MEA): Study the effects of operating conditions on the formation of non-regenerable compounds, on excessive solvent losses, on corrosion and on foaming

- Diisopropanolamine (DIPA): Study effect of operating pressure on hydrogen sulfide removal efficiency

- Diglycolamine (DGA): Study the effects of operating conditions on the formation of non-regenerable compounds and the effect of feed acid gas composition on removal efficiency at various operating temperatures and pressures

- Diethanolamine (DEA): Study means of removal of the fine particles that cause foaming, as removal efficiency vs. operability vs. cost. Data are needed on utilities requirements vs. operating temperature and pressure

- Fluor Solvent: Determine utility requirements and study effect of feed gas composition on removal efficiency at various operating temperatures and pressures

- Sulfinol: Determine solubility of hydrocarbons in the sulfinol solvent and study process economics vs. operating parameters

- Estasolvan: Study the effect of operating pressure on acid gas removal efficiency. Study methods of treatment for the blowdown stream

- Benfield: Study the process when it is operated selectively for the hydrogen sulfide content in the carbon dioxide stream. Determine the extent of COS hydrolysis vs. the requirements for Stretford process feed

- Amisol: Determine utility requirements and study effect of feed gas composition on removal efficiency at various operating temperatures and pressures

Sulfur Recovery and Tail Gas Cleanup

Further study of the sulfur recovery and tail gas cleanup process should include characterization of inlet and outlet gas streams, vent streams, byproducts, sulfur removal efficiency vs. operating parameters, and reactant degradation. Suggested areas of investigation in examples of processes include:

- Claus process: Determine extent of removal of HCN and ammonia from the feed gas stream and fate of CO and hydrocarbons in the feed gas. Study the effect on sulfur conversion of the presence of oxidizable compounds in the feed gas. Determine the economics of the process for operation on various feed gas compositions, with particular emphasis on the effects of variations in hydrogen sulfide concentration

- Stretford and Beavon processes: Determine conversion of organic sulfur compounds in the presence of high concentrations of carbon dioxide. Characterize oxidizer vent gas stream and solvent blowdown stream. Determine degree of removal of mercaptans and ammonia

Sulfur Dioxide Control Processes

For the citrate process, data are needed on the effects on sulfur dioxide removal and on the process economics of changes in feed gas composition.* These same data are needed for the Kellogg-Weir scrubber. For the Wellman-Lord process, data are needed on the efficiency of removal of HCN and ammonia.

For all processes, data and information are needed to supplement those already available on composition and means of disposal of the products from each process, such as sulfur, sulfuric acid, concentrated sulfur dioxide,and sludges.

These examples are cited to illustrate the types of data that are needed for process and economic evaluation of sulfur dioxide control processes.

Hydrocarbon Control Processes

The hydrocarbon content of waste gas streams can be reduced or eliminated by incineration, by absorption,or by adsorption.

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*R. S. Madenburg of the Morrison-Knudson Company, Boise, Idaho, presented a paper, "Industrial Application of Citrate FGD Technology," at the June, 1978 meeting of the Air Pollution Control Association, indicating the areas of use of the process, its efficiency and costs. More data are needed, however, on process variations such as feeding dilute hydrogen sulfide at elevated pressure in order to maintain a partial pressure of hydrogen sulfide at 10 to 14 psi.

Incineration by oxidation at high temperature is attractive as a means of hydrocarbon destruction because the waste gas stream can be introduced directly into the utility boiler. Boiler design is affected by the quantity and composition of the waste gas streams and data are needed on the economic effects of these design changes.

Absorption into the solvent of low vapor pressure followed by stripping and solvent regeneration has been proposed. More data are need on process types, efficiencies and economics as these are affected by stream composition and quantity.

Carbon adsorption is effective for removal of some hydrocarbons and, like absorption, yields a concentrated hydrocarbon stream upon regeneration. Data are needed on efficiencies and economics vs. gas stream composition and quantity.

Nitrogen Oxides Control Processes

Nitrogen oxides emissions can be controlled by combustion modification in boilers and furnaces, by using fluidized bed combustion techniques, by hydrodenitrogenation of liquid fuels, and by flue gas cleaning. The available data on these methods are drawn from power plant boiler experience and petroleum refinery experience. Data are needed on application of the techniques to the combustion problems peculiar to coal conversion plants, particularly where the utility boiler may also act as an incinerator for waste process streams. These data should include the effects on economics of the installation of the control methods at various levels of control efficiency. Investigations might be started in pilot plants and later be extended to full scale plants.

As an example of the types of data needed, suggested areas of investigation into fluidized bed combustion (FBC) are as follows:

- Determine the sulfur retention in the ash from FBC and spreader-stoker boilers for representative coals and mixed fuels
- Determine nitrogen oxides emissions for the same boilers and fuels
- Determine the fate of trace elements, including flue gas desulfurization systems
- Develop economics as a spur to development of the process

For UOP-Shell, Hitachi-Zosen and other processes more investigation is needed into the effect on the processes of particulates from coal fired boilers.

More data are needed on nitrogen oxides formation in boilers with and without combustion modification in order to determine the benefits of the modifications. Development of cost data is needed.

For liquid fuel denitrogenation, data are needed on the nitrogen content and type in coal derived liquids. Study of denitrogenation of these liquids is needed, together with development of economic data.

Exxon Research and Engineering Company is developing and applying a system described in "Reducing Nitrogen Oxides Emissions by Ammonia Injection," a paper given by R. K. Lyon at the June, 1978 meeting of the Air Pollution Control Association. The described system achieves reductions of around 70 percent by injecting ammonia into the boiler flue gas. There have been 8 demonstrations in commercial boilers and furnaces, to date. Further development of applications, efficiencies, and costs for this simple process should be encouraged.

Cooling Tower Drift

More technical and economic studies are required to develop more data on means of reducing cooling tower drift. Economic projections would be useful in contrasting high velocity cooling tower designs with low velocity and hyperbolic designs.

Lock Hopper Vent Gas

Study is needed to determine the optimum type or types of lock hopper operation and methods of disposal of excess lock hopper vent gas in the various coal conversion processes.

Developent of extrusion type coal feeders for high pressure reactors is progressing. Use of these would reduce or eliminate vent gases. Inclusion of such process wastes as tars, oils and biological oxidation sludges with the coal feed might aid the operation of the feeders and incidentally solve a disposal problem.

SECTION 10

ENVIRONMENTAL DATA ACQUISITION: CONTROL OF SOLID WASTES

The many types of solids with widely varying chemical and physical characteristics that are involved in the operations of coal conversion plants pose a variety of problems concerned with:

Fugitive dusts from:
- Coal storage and reclaiming operations
- Coal storage piles
- Ash and scrubber sludge piles
- Limestone storage piles
- Sulfur storage, handling, and shipment

Bulk handling of:
- Coal
- Coal fines and dust
- Conversion process ash slurry
- Incinerator/Boiler ash
- Scrubber sludge
- Spent catalysts
- Evaporated solids
- Limestone
- Sulfur

Disposal of: Coal fines and dust
Ash
Sludges
Spent catalysts
Evaporated solids

These many problems may be broadly grouped into the general categories of dust control and solid waste management. Efforts in both categories are directed toward minimizing or eliminating deterioration of the environment. In the subsections that follow, the overall problems are examined and controls are proposed to meet existing and proposed environmental standards. Developing control technology is examined in view of future environmental goals. In all cases the viewpoint of the conversion plant operator is taken: solutions for environmental problems should be realistic, operable,and economical.

LITERATURE SURVEY AND DATA GATHERING

In the survey of available literature on coal conversion and related processes data and information were gathered on the evolution of fugitive dusts, methods of control that are in general use, suggested means of more efficient dust control, the extent to which solids disposal problems are recognized, proposed means of solving solids disposal problems and possible future development of more efficient or lower cost solids management methods. Close collaboration was maintained with the effluent and emission study groups in this project for recognition and solution of problems that required consideration of more than one state of matter.

Personal Contacts, Trips, and Meetings

In addition to the list of contacts and meetings that was given in Section 5 of this report, there were others that yielded information on solids handling:

- A visit to the Public Service Company in San Antonio, Texas, to observe coal receiving and stockpiling operations including dust suppression, reclaiming, and handling of dust and fines
- A meeting with Dravo Corp. engineers for a general discussion of coal and dust handling and a particular discussion of the Bi-Gas pilot plant at Homer City, Pa. Visits to the pilot plant, a coal fired power generating station and U. S. Steel's coal preparation plant in Kirby, Pa.
- Attendance at the Coal Waste Technology Seminar in Houston.

Telephone contacts (TC) and correspondence (C) were held with the following:

- Dravo Corp. - F. A. Zuhl (TC). Information on "Calcilox" waste solidification system
- Chemfix Inc. - Alan Cohen, Richard Patent (TC). Information on I.U. Conversion Systems solidification process
- McDowell-Wellman Engineering Co. - J. Wellman (TC). Dust control at rail car dumping stations

- Allis Chalmers Co. - B. K. Smay (TC). Dust control
- EPA Office of Solid Waste Management Programs - Allan Geswein and Robert Landreth (TC, C). Liners for land disposal sites
- Austin (TX) Water Control Board - J. Snow, J. Carmichael, G. Maxon (TC, C). Ash evaporation ponds and liners
- DuPont Inc. - Phillip Rizzo (TC), C. S. Glover (C). Information on pond liners
- American Colloid Co. - William Hahn (TC, C). Bentonite in waste disposal areas
- Dowell Division of Dow Chemical Co. - Dan Hunt (TC, C). Soil sealant
- B. F. Goodrich General Products Co. - H. F. Cummings (C). Hypalon liners
- National Ash Association, Washington, D. C. (TC). Disposal pond operation
- Johnson-March Co. - C. L. Burchsted (TC, C). Dust suppression

TARGET POLLUTANT RESIDUALS

Guidelines and standards of federal, state, regional, and international jurisdictions for solid waste disposal requirements were reviewed and synopsized in the same manner as were those for air emissions and liquid effluents.

Solid Waste Disposal and Management

The majority of the solid waste disposal requirements are much less definitive, with regard to establishing design requirements, than those criteria established within the air and water regulatory areas. The regulations tend to establish requirements directed more toward the operation of a disposal facility, such as adequate rodent control and proper compaction and cover for solid waste, than to the design.

It should be expected that the regulatory activity in this area and especially with respect to hazardous wastes will continue to increase as a result of the Solid Waste Disposal Act as amended by the Resource Conservation and Recovery Act of 1976, Title II Solid Waste Disposal (42 USC 6901 et seq.).

One provision generally common to the states reviewed allows for solid waste disposal on one's own property without a permit so long as no nuisance conditions are created.

Texas, one of the states surveyed, has issued Technical Guidelines for solid waste disposal and indicates that by following these guidelines all solid waste disposal requirements will be satisfied. These Technical Guidelines are available from the Texas Water Quality Board, which has responsibility in this area.

In "Guidelines for the Land Disposal of Solid Wastes" (40CFR241) is the EPA requirement that location, design, construction, and operation of land disposal sites shall conform to the most stringent of applicable water quality standards. States require leachate collection and treatment and may require water monitoring (N. Dak.) or leaching and drainage are to be prevented (Ky., Mont., Wy.).

The Dominion of Canada sets criteria for impermeable soils for disposal sites:

- Permeability to be not more than 10^{-6} cm/sec
- Not less than 30 percent passing U.S. standard 200 mesh sieve
- Interceptor drains or collectors may be required

Other most stringent landfill standards and guidelines applicable to ash and inorganic sludges are as follows:

- Dust control measures are required (N. Dak.)
- Cover the fill site at least every 20 days (Br. Col.)
- Slope the cover 2 to 4 percent (Can. Fed.)
- Neutralize and dewater sludges prior to storage (Br. Col.)
- Compaction is required (Alaska, Pa.)
- Closure with 24 inches of compacted earth (N. Dak.)
- Surface drainage to be consistent with surrounding area, not to cause interference with adjacent drainage nor allow runoff to become concentrated (N. Dak.)
- Area to be seeded after closure (N. Dak.)
- No solid waste is to be discharged into ground or surface waters (Tex.)

Pennsylvania has regulations applying to coal mining refuse piles which presumably could be used as solid waste disposal guidelines:

- Maximum height of the pile to be 100 feet above average elevation of the immediate area
- Refuse is to be deposited in layers no more than 2 feet thick and is to be compacted
- Surface water is to be managed (in a manner similar to that described previously).

Fugitive Dusts

Fugitive dust standards and guidelines are similar to those for solid waste disposal in that there are few numerical values and much subjective direction:

- "No material shall be handled, transported, stored, or disposed of....without taking reasonable precautions to prevent particulates from becoming airborne." Use of water, asphalt, or oil emulsions to control dust is suggested (N. Mex., Ala. and others)
- "Coal preparation plants. All crushers, conveyors, screens, cleaners, hoppers, and chutes, which are designed for continuous transportation or preparation of coal shall be equipped with hoods, shields, or sprays where reasonably necessary to prevent airborne particulate matter" (N.Mex.)
- "....without taking reasonable precuations to prevent particulate matter from becoming airborne in amounts which cause a public nuisance" (N. Dak., Okla. Va., Wash.)

- "....without taking reasonable precautions to prevent particulate matter from becoming airborne" (O., Tenn.)
- "Any source must have control equipment to minimize emissions" (W. Va.). Applies to ash or fuel piles, transport and handling
- Precautions are to be taken to control fugitive dusts through use of water, chemicals or covering (Texas)

There are a few numerical values stated:

- Fugitive dusts shall not exceed 20 percent opacity except when the wind exceeds 30 MPH (Colo.)
- No particles larger than 40 microns mean diameter shall be airborne beyond the property line except when the wind exceeds 25 MPH (Ill.)
- Maximum allowable ground level concentration at the property line is 2 milligrams per cubic meter above such ground concentrations for periods of not more than 10 minutes in any hour (Kan.)
- Cannot exceed ambient air standards (Wy.)
- Visibly passing property lines is prohibited (Colo., Okla.,and others)

Odors

With the exception of Missouri, West Virginia, and Wyoming, which describe apparatus or evaluation methods, the statements concerning odors are all subjective, as "malodorous air shall not be emitted such that odors are detected beyond the property line," "no odor that would be objectionable to a person of ordinary sensibility shall be emitted," "best available control technology shall be used," "best practical control technology shall be used," "prohibited to allow emission of odorous gases, liquids or solids in such quantities as to cause air pollution."

Consideration of these regulations leads to certain conclusions regarding odor control in solids handling and waste disposal in caol conversion plants:

- o Spontaneous combustion in coal storage piles should be avoided to avoid emissions of hydrogen sulfide, sulfur dioxide, and other odorous materials
- o Organic materials should not be allowed in ash and ash/sludge disposal areas
- o pH control of the liquid portion of the ash slurry to disposal may be necessary to prevent odor-causing reactions within the disposal area

DUST CONTROL

Suspension of finely divided solids in ambient air as fugitive dusts is governed by the size, shape, and density of the solids and by the air velocity and direction. In air suspension or by settling in areas surrounding coal conversion plants the dusts can become a fire hazard (in the case of coal and sulfur), a physical hazard, a contributor to deterioration of the environment by discouraging plant growth or a toxic factor in the environment (through inhalation by man and animals or by the leaching out of soluble compounds into adjacent land and water). Control of these dusts is important from the environmental standpoint and from the economic considerations of loss of usable materials or of loss of useful land.

Fugitive dust clouds are generated when a stream of dry solids falls freely in air as at a conveyor transfer or discharge point, when the stream of solids is agitated as at rail car dumping or reclaiming from pile storage and when wind blows across piles of solids. There is little quantitative data on fugitive dust generation. Available information is principally concerned with gross material losses and not with measurement of cause versus effect. Process operators, however, are making efforts to reduce dust evolution from materials handling operations and, aided by equipment manufacturers, have developed systems for dust control that may be characterized as dust suppression and dust collection.

Definition of the Problem

Run-of-mine coal is crushed and screened during the preparation operations for removal of inerts and pyrite-bearing particles. The size distribution of the crushed particles is affected by:

- Coal source and type
- Crusher type
- Crusher design
- Crusher discharge opening
- Crusher speed
- Coal grindability

The effect of these variables on the size distribution of the crushed particles is illustrated in TABLE 10-1 for crushers designed by two manufacturers operating on several types of coal.

TABLE 10-2 contrasts the operation of these two crushers and a third model all operating with 1.25 inch crusher opening on various coals. The size data are shown in graphical form in Figure 10-1.

Other data on size consists is reported for Indiana and Illinois coals (2) and for "western" and "eastern" coals (3) and is shown in TABLE 10-3. Unfortunately, the crusher size opening was not specified and comparison with the data in TABLES 10-1 and 10-2

TABLE 10-1. SIZE DISTRIBUTION OF PRODUCTS FROM COAL CRUSHING

CRUSHER A* ON ILLINOIS NO. 6 COAL, % RETAINED ON SCREENS

Screen Opening	Crusher Opening, In. (S=Slow Speed, F=Fast Speed = 1.65 to 1.95S)										
	2.00S	1.25S	1.00S	0.75S	0.50S	0.25S	1.25 F	1.00F	0.75F	0.50F	0.25F
1.50 in.	8										
1.25	15	4.5	1.0				3.1				
1.00	30	12.9	5.3	2.0			9.4	2.8	0.2		
0.75	49	28.9	15.1	4.5			25.1	11.3	6.2	1.2	
0.50	68	49.7	38.0	32.2	2.9		45.6	34.1	23.3	4.4	
0.375	75	61.4	50.4	45.7	16.0		57.8	45.9	41.3	10.2	
0.25	84	72.8	65.4	61.9	26.5	1.2	68.2	61.5	57.4	31.7	2.4
0.125	91	82.9	80.6	77.2	57.1	30.5	78.7	76.4	72.8	52.2	26.3
10M**	95	87.0	84.3	82.7	71.2	47.8	83.3	80.2	77.6	65.5	43.6
28M	98	93.0	92.1	90.9	82.9	74.5	89.8	87.8	85.4	78.3	70.3

CRUSHER B* ON VARIOUS COALS, % RETAINED ON SCREENS

Screen Opening,	Crusher. Opening, In.	W. Ky. 1.50	W. Ky. 1.00	Ind. 1.00	Ill. 1.25	Ill. 0.75	Ohio 2.0	Ohio 1.50	Ohio 1.25	W. Va. 1.50	W. Va. 1.25	W. Va. 1.00	Wy. 2.0
2.00 in.							0.6						5.6
1.75													10.7
1.50		2.7					7.1			3.6			
1.25		12.6			5.5			3.0	0.5	28.1	8.8		27.5
1.00		31.9	6.8	5.4	24.0		51.0q	17.3	11.2	47.9	26.9	2.1	37.3
0.75		54.3	34.8	31.3	52.3	11.0	65.5	44.5	35.4	68.5	55.9	20.6	48.9
0.50		71.1	58.4	61.3	73.1	50.5	77.2	67.8	61.2	80.2	73.4	51.6	62.1
0.375				71.5									69.3
0.25		85.3	77.9	81.6	88.5	80.9	88.2	83.2	79.3	91.6	88.3	78.4	77.6
0.125									88.6				85.7
10M**			88.2	91.9	94.9	91.9							
28M			95.2										

* Crusher manufacturers' names confidential, by request
** Tyler standard mesh size designation

TABLE 10-2. COAL CRUSHING WITH 1.25-IN. OPENING

(Percent Retained on Screens)

Screen	Illinois		W. Ky.		Ohio		W. Va.
Size	A*	B*	C*	B	C	B	B
1.25 in.	4.5	5.5	7.2	3.9	23.0	0.5	8.8
1.00	12.9	24.0	22.5	17.9	35.7	11.2	26.9
0.75	28.9	52.3	46.8	44.5	49.5	35.4	55.9
0.50	49.7	73.1	66.1	65.4	64.9	61.2	73.4
0.25	72.8	88.5	82.0	81.9	79.7	79.3	88.3
0.125	82.9				87.0	88.6	
10M**	87.0	94.9					
28M**	93.0						

* Crusher manufacturers' names confidential, by request

** Tyler standard mesh size designation

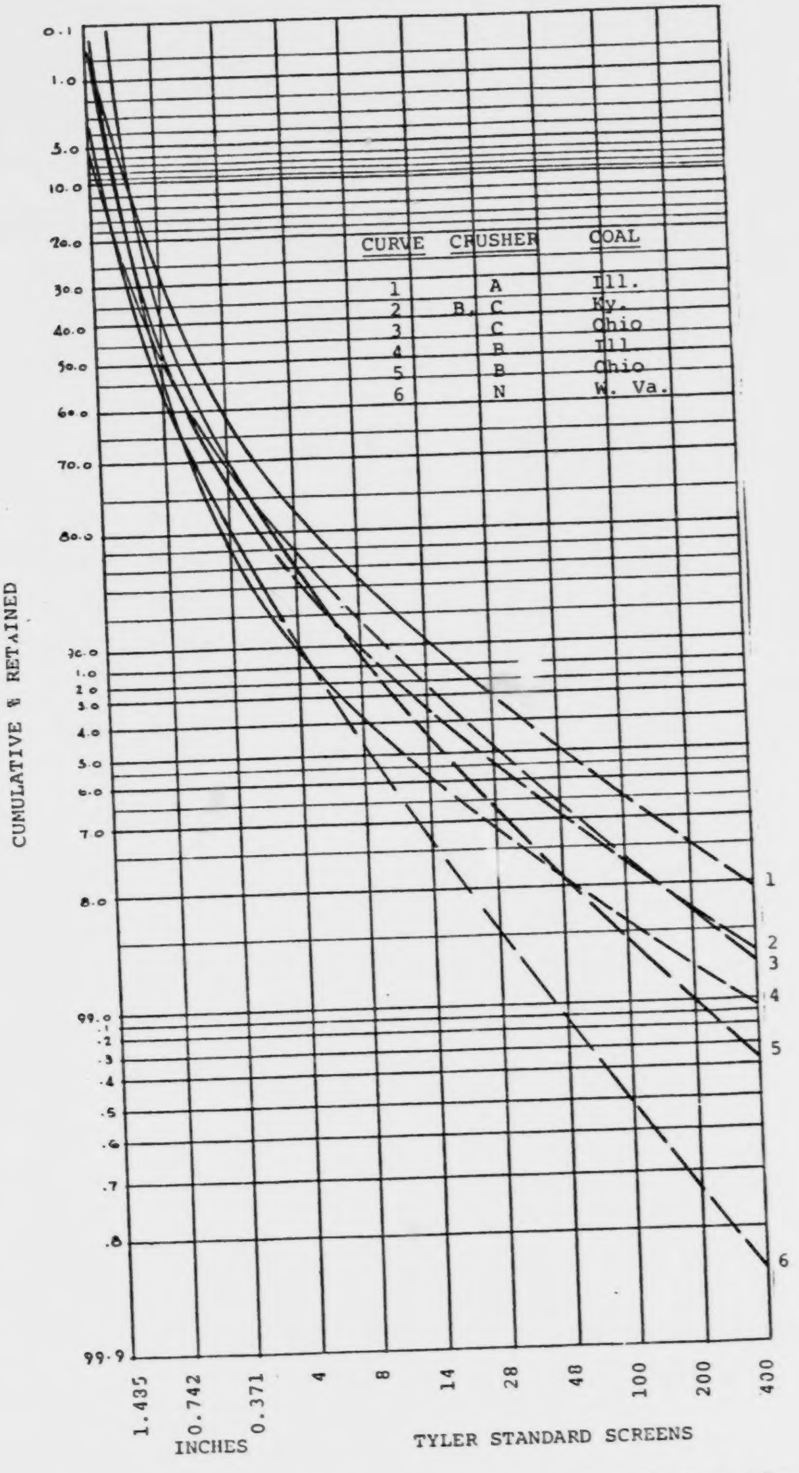

Figure 10-1. Size Consist of crushed coals.

TABLE 10-3. SIZE CONSIST OF AS-RECEIVED COALS

(Percent Retained on Screens)

	A*	B	C	D	E	F
1.500 in.	3.0	3.0		2.0		
1.250	12.0					1.0
1.000	21.0	15.5	12.0	8.5	3.3	
0.750			27.9		12.9	14.0
0.500	45.0	44.5	47.0	24.0	34.5	
0.375					43.6	41.0
0.250	61.5	57.5	71.4	43.0	63.4	
0.125			87.3			74.0
10M**	83.5	74.0		74.0	80.9	
28M		87.0		88.0		90.0
100M					97.7	
200M					98.7	

* A = Old Ben No. 1, Indiana Strip Mine Washed (2)
B = Old Ben No. 24, Illinois Raw Coal (2)
C = Consolidation Norris, Illinois Strip Mine (2)
D = Old Ben No. 26 , Illinois Deep Mine (2)
E = Western Coal (3)
F = Eastern Coal (3)
** Tyler Standard Screen Mesh

can be made only by inference.

The impact of fugitive dust from coal storage and handling was assessed (1, p.IV-16ff.) by comparing coal piles to piles of crushed stone and to road and soil dust. The conclusions reached in the study may be applied to coal conversion plants, since the operations are very similar.

The study concluded that because coal in a storage pile would not purposely be pulverized until just before injection into a coal conversion plant, it is reasonable to assume that only a small portion of the particles is capable of wind transport over long ranges; in general, only particles of diameter less than 30 micrometers have long-range drift potential, i.e., greater than 1,000 feet from origin.

By comparison, dust normally encountered from unpaved roadways, construction sites, and agricultural soil operations contains from 65 to 85 percent of particles that have diameters smaller than 30 micrometers. Of these, over half are smaller than 2 micrometers. Particles in this latter category have drift potential of hundreds of miles. Therefore it may be reasoned that dust generated from coal storage piles and coal handling generally has localized (1,000 feet) impact, while dust generated from such sources as construction sites, farming activity, and unpaved roadway travel has potential for long-range air quality impacts.

Although considerable uncertainty is introduced when the curves of Figure 10-1 are extrapolated to 400 Tyler mesh (37 micrometers), the inference may be drawn that the content of particles in crushed coal less than 400 mesh appears to be in the range of 0.2 to 2 percent and may average about 1 percent. From the foregoing study, this amount of coal dust represents the maximum amount of dust that may drift beyond property lines and be considered in excess of environmental standards.

The conclusions of the assessment study do not address the problems of fugitive dust within the plant boundaries, however, and these, because of the potential fire and toxic hazards, must be considered in assessing the overall problem.

The relationship between particle size and distance that a particle will be carried may be demonstrated by use of the following equation, assuming Stokes' law is applicable:

$$D = [(18\ uHV)/(gLd)]^{0.5}$$

D = minimum particle diameter collected
u = air viscosity
H = vertical distance (pile height)
V = air velocity
g = local acceleration of gravity
L = horizontal travel distance
d = (density of particle) - (density of air)

Thus, for a 30 micrometer particle to reach the ground 1,000 feet from the peak of a coal pile 45 feet high, the crosswind velocity can be no higher than about 2 miles per hour.

Distance from the coal pile and wind velocity have a profound effect on the diameter of the smallest particle that will reach the ground from the peak of a 45 foot pile:

	Crosswind Velocity			
Distance from peak, ft.	2 mph	5 mph	10 mph	20 mph
100	95(170)*	150(100)	212(65)	300(48)
500	42(325)	66(200)	94(170)	133(115)
1,000	30(575)	47(300)	67(200)	95(170)

*Particle diameter in micrometers. Approximate equivalent standard Tyler screen mesh shown in parentheses.

From Figure 10-1, on the order of 0.5 to 4 percent of the coal could reach the ground 500 feet from the pile with a 20 MPH crosswind. In a gasification plant of 250 billion Btu/day capacity, this potential dust accumulation could be 75 to 700 tons per day, depending on the coal source and the conversion process.

Another investigation determined the minimum size of particles that could safely be exposed to wind action (7), with the following results:

Particle Diameter Micrometers	Tyler Screen Opening	Wind Susceptibility
0 to 420	< 35	Highly erodible
420 to 840	35 to 20	Difficultly erodible
840 and over	> 20	Usually non-erodible

Dust Suppression by Water Sprays

The usual operation at the coal unloading station in power plants involves dumping cars by inversion into a receiving hopper and conveying the coal to a stacker for distribution in the storage area. All of these operations are potential producers of fugitive dust.

Dust at the car dumping station is usually controlled by enclosing the area and directing water sprays into the car and into the receiving hopper to wet the coal particles at the moment of discharge so that the small particles agglomerate and do not become airborne. Surfactants may be added to the spray water to increase the wetting action, most frequently at a dilution of about one part surfactant to 1,000 parts water. Solution application rate averages about 2 gallons per ton of coal, increasing for dry coal and decreasing for wet coal.

When wetting agents are used there is a carryover effect to the storage pile that aids in reducing dust generation at the discharge of the stacker to outside storage piles. Additional sprays may be used during reclaiming operations to ensure that all coal is wetted.

When plain water is used in the spraying system, on the order of 5 to 8 percent of moisture, or about 12 to 20 gallons of water per ton of coal, must be applied, in contrast to the (approximately) 0.8 percent moisture addition at 2 gallons per ton when surfactants are used.

Design of the wetting system is critical, since the coal is discharged from the car by a rotary dumper in about 30 seconds. The principal objectives are (5):

- o Confinement of the dust within the dust producing area by a curtain of moisture
- o Wetting the dust by direct contact between the particles and water
- o Formation of agglomerates too heavy to remain airborne, or too heavy to become airborne, by encouraging adhesion of particles to each other or to larger coal particles with water as the adhesive

Careful attention to formulation of the surfactant yields a wetting solution whose surface tension may be about 35 to 40 percent of that of water. The effects of the surface tension decrease are (5):

- o Dust can pentrate into the water droplets, rather than only coating the surfaces. The entire volume of the droplet is available for dust collection
- o Fine particles, because they are more readily wetted, are cemented to each other and are effectively encapsulated by the surface active droplet

- The aggregates, because of increased weight, either drop from the air stream or cannot become airborne
- Sprays of untreated water yield large droplets with low surface area and large, wasted volumes. Treated water, because of the reduction in surface tension, is more readily atomized, more droplets are produced per unit volume, total surface area of the droplets is greatly increased and the contact potential with dust particles is vastly improved.

The surfactant selected for use should exhibit such characteristics as the following (5):

- Excellent surface tension reduction for optimum wetting, spreading, and penetration
- Heat stability, freeze-thaw stability
- Miscible with water in all proportions
- No flash or fire point
- Non-toxic
- Non-corrosive
- Contain biocidal additives to inhibit growths of slimes that may clog sprays, strainers, or parts of the proportioning system

Completely engineered systems for dust suppression by spraying of water solutions of surfactants are available from specialists in this field. Observation of systems in operation when receiving coals that are already wet or that contain normal (6 to 10 percent) moisture led to the following conclusions:

- Dust at the car dumping station is greatly reduced by the sprays
- Dust is evolved, despite the sprays, in quantities sufficient to constitute a nuisance to personnel in the enclosed or semi-enclosed dumping areas and to constitute a possible potential explosion or fire hazard from dry dust

settling on surfaces of the dumping station

- Periodic washdowns of the dumping area are needed to control buildup of settled dust
- Dust suppression at the storage pile stacker is effective for most coals
- The sprayed-on material evaporates rapidly enough in live storage to lose much of its effectiveness with the result that wind-generated fugitive dust from the pile remains a problem

In general, the water-plus-surfactant spraying systems have greatly reduced the fugitive dust problem, but have not eliminated it. This is especially apparent when dry coals are received and the deficiencies of the control system are more readily visible. Much of the difficulty seems to lie in the necessarily short exposure to sprays of the coal stream dumping from the cars: while most of the surfaces of the larger coal particles are wetted, much of the dust that becomes airborne during dumping escapes contact with the water droplets and is not collected. This situation might be improved by installation of more sprays of the misting type at the dumping station. A likelier solution may well be the use of a more dilute solution of surfactant, for example 1:4000, in flooding type sprays directed downward into the cars before they are dumped to more nearly saturate the coal and cause dust particles to adhere to the larger particles or to agglomerate. Excess water draining from the cars could be recycled through strainers to the flooding sprays.

Costs have been published (mid 1974) for a spray system installed in a 1,000 TPH rock crushing plant to spray 1.5 gallons of wetting agent solution per ton of rock processed (6). Installed equipment cost was estimated at $61,676. Operating costs were estimated on the basis of 1,920 hours per year plant operation (one shift at 8 hours per day, 5 days per week, 48 weeks per year) with spray systems operation assumed to be required only 40 percent of the plant operating time due to the moisture content of the material or the prevailing weather conditions. Total operating costs on this basis were reported as $17,050 per year, or $0.009 per ton of stone produced.

These costs will be treated later in this report for application to coal conversion plants.

Dust Suppression by Total Wetting

As an alternate to spraying coal with surfactant solutions, consideration might be given to totally wetting the coal prior to unloading. In this case no surfactant would be used, the whole carload of coal would be wet and maximum agglomeration would be achieved.

Bituminous coal as received at the conversion plant will contain on the order of 10 percent moisture, subbituminous on the order of 15 to 25 percent and lignite on the order of 25 to 35 percent. This moisture is the water remaining after processing in the coal preparation plant, handling, storage, and transport to the conversion plant site. The amount of water that can be held by the coal is partly a function of the overall particle size distribution and partly a function of such particle surface characteristics as porosity and the tendency to be hydrophilic or hydrophobic. There are no actual data on the water-holding characteristics of various coals, probably because of the many variables involved. Consideration of the following general statements from Davis Associates' "Coal Preparation Environmental Engineering Manual" (4) may aid in estimating the total water content:

1. Natural drainage will reduce surface moisture to 5 to 6 percent for particle sizes larger than 0.5 inch.
2. Vibrating screens reduce moisture to about 10 percent on particles coarser than about 0.75 inch.

3. Finer fractions can be expected to contain on the order of 20 to 40 percent moisture after screening, centrifuging or filtration.

4. The natural moisture content for coarse bituminous coal refuse (median about 10mm) is about 11 percent and for fine refuse (median about 0.1mm) is about 31 percent.

5. An example is given (p.330) for a typical coal cleaning plant producing:

Coarse Coal	270 tons		6% moisture (approx.)
Intermediate coal	363		10%
Fine coal	64		30%
	697 tons	Ave.	10% moisture
Coarse refuse	190 tons		11% moisture (approx.)
Fine refuse	113		31%
	303 tons	Ave.	18.5% moisture

The average coal as shipped contains 0.11 pounds of water per pound of dry solids while the average refuse as piled in disposal areas naturally contains 0.23 pounds of water per pound of dry solids. Recognizing that the refuse particle sizes are generally smaller than the coal sizes and that the water holding capacity of the refuse is therefore greater than that of the coal, the natural (saturation) water content of the coal may be estimated to be an average of about 0.17 pounds of water per pound of dry solids, or an increase of about 50 percent over the shipped moisture content.

Applying this reasoning to three coals received at a conversion plant producing 250 billion Btu/day of synthetic natural gas yields the following water requirements for saturating all coal received:

	Bituminous	Subbituminous	Lignite
Dry coal feed	17,000 TPD	17,000 TPD	17,000 TPD
Moisture as received - wt%	10	20	30
- lb/lb dry solids	0.11	0.25	0.43
Saturation, lb/lb solids	0.17	0.37	0.64
Moisture gain, lb/lb solids	0.06	0.12	0.21
Water required-gal/day	245,000	490,000	858,000
As received coal	18,900	20,400	22,100
Cars per day, 100T capacity	189	204	221
Water per car, gal.	1,300	2,400	3,900

The operating concept is: flood the car just before dumping over a collecting basin; drain the basin, the dump hopper, the conveyor feeders, and the conveyor tunnel to a sump; recycle the water to the flooding nozzles. Since the nozzles are large, only a strainer would be needed to remove large particles from the water.

Dust Suppression by Chemical Binders

The stockpile of feed coal for conversion plants is usually considered to be about 14 days' feed, or about 300,000 tons. Site specific conditions, such as the logistics of coal movement, may require considerably more stockpiling. Reclaiming from stockpiles is usually accomplished by conveyors in a tunnel or tunnels beneath the stockpile, fed to a series of openings.

The volume of coal directly over the openings flows to the conveyors by gravity to form conical depressions with walls sloping roughly 40 to 45 degrees from the horizontal. This center, or "live", storage volume moves through the storage area in a relatively short time and, if a surfactant solution is well distributed on the coal during unloading and stacking, will contribute little to the fugitive dust problem because of the residual effect of the solution. Depending on the stacking method and the number of openings under the total storage pile, the live storage volume may vary from 20 percent of the total volume (one pile, one opening) to 55 percent of the total (multipile openings) (4, p.280).

The remaining 45 to 80 percent of the total volume of coal stored is moved only when the live storage volume is exhausted. This "dead" storage volume may therefore be immobile for considerable periods of time, any carryover effect of surfactants from the unloading sprays is soon lost and the pile may become a source of fugitive dust.

Chemical binders of several types are available for application to the outside of dead storage piles to alleviate the dusting problem (7). These chemicals coat the topmost particles with a thin film, promoting adherence of particles to each other, and forming a tough, durable crust that is resistant to wind and rain erosion but through which moisture can penetrate and run in the void spaces between particles. While the crust is intact the pile is protected from windage loss and erosion from rainfall and the dust problem is eliminated. Once dried, the film is insoluble and tough enough to be resistant to the strains of freezing and thawing. Examples of chemical binders are:

- Compound SP (Johnson-March Corp.) is a water base synthetic organic polymer that will coat all types of materials and sizes of materials and dries to an inert, non-toxic film that does not affect coal burning qualities. Various grades are made for various conditions. Application rate is 1 gallon per 100 square feet of pile surface.

- Coherex (Golden Bear Oil Co.) is a mixture of semi-liquid petroleum resins and a wetting liquid containing sequestering agents. The resins form films and bind dust while the wetting liquid carries the resins, wets the particle surfaces and renders the compound miscible with water in all proportions. The mixture is non-flammable. Usual application rate is about 1.5 gallons of Coherex per 100 square feet of pile surface. The Coherex is diluted with 4 parts of water prior to application. Information for Coherex (7) shows a cost of $0.225 per gallon for the material.

- Dowell Binder (Dowell Division of the Dow Chemical Co.) is a specially formulated synthetic liquid adhesive that is diluted to about 1 part in 24 parts water for spraying. On drying a crust about 1.5 to 2 inches thick is formed that eliminates wind and rain erosion from coal in transit as well as coal in stockpiles.

- Aerospray Binders (American Cyanamid Co.) are polymeric (Aerospray 70) and alkyd resin (Aerospray 52) water-dispersable materials, the former for short term use, the latter for long term. Both Aerosprays form surface films when particles are small and dilution is small and penetrate when particles are larger and dilution is greater. Usual dilution for Aerospray 70 is about 1:4 with application of about 5 to 6 gallons of solution per 100 square feet. For Aerospray 52 the dilution is about 1:10 with application of about 3 to 4 gallons per 100 square feet.

Dust Suppression by Physical Binders

Stockpiles may be coated with asphalt or road tar to provide an airtight seal. The procedure is quoted (7) from the "BCI Fuel Engineering Data Book", Section D-3, prepared by the Bituminous Coal Institute:

> "In this method the top and sides of the stockpile are covered with an air-tight seal of asphalt or road tar. This seal may be applied with spraying equipment similar to that used by highway departments. The top of a large storage pile can be capped by the use of a hose and hand nozzle. Small piles can be capped by hand spraying only.
>
> A covering or cap, 1/8 in. thick and requiring about 1 gal. of asphalt per 10 sq ft of area, has been found satisfactory. The sides exposed to prevailing winds should be treated somewhat more heavily for greater assurance against access of air.
>
> A preference has been shown for the AE-4 grade of asphalt emulsion because of its quick water-separation and superior coating properties."

Dust Suppression by Compaction

In another quotation from the "BCI Fuel Engineering Data Book" (7) are given recommendations for compaction sealing stockpiles against weather. The sealing method has the additional advantage that wind erosion is minimized:

> "Plants having reserve piles larger than 500 tons should always build a compacted coal pile. Compaction seals out air and minimizes spontaneous heating. It also reduces heating-value loss to one percent or less.
>
> Dumping coal on the pile segregates it, creates flues and promotes heating and fires. The pile should be built in successive layers 1 to 2 ft thick. Each layer should be thoroughly compacted by repeatedly running a bulldozer or weighted roller over the coal.
>
> Dress the pile by sloping the tops of successive (coal) layers and compacting the pile toward the sides. The sides of the pile must be compacted and dressed. This prevents rain, melting snow and ice from penetrating the pile. Pile slopes should not be too steep; and angle of 30 deg to the horizontal is a good slope to maintain. Steep side slopes can cause segregation and erosion from wind and rain.
>
> Crown the top of the pile slightly or allow for slope in one direction. Saucer-shaped tops create a problem of standing water, wetting the coal so that it clogs chutes and conveyors.
>
> It is recommended that, insofar as possible, only one type of coal be placed in the stockpile. For large compacted reserve storage piles, the coal should be a slack or nut and slack size not exceeding 2 in. top size. Size breakdown should be such as to fill all the voids in the coal as it is compacted. Large size or double-screened coal should never be compacted.
>
> A 6-in. layer of fine coal (1/4-in. x 0) [should] be placed over the top and slopes of the shaped stockpile. As an anchor, place over this a 4-in. layer of larger size material (at least 2 in. x 0). The combination will materially prevent wind and rain erosion."

Dust Reduction by Agglomeration--A Concept

As an alternate to dust suppression, if dust were removed from coal at the mine, agglomerated and then recombined with the coal before it is loaded into cars, most of the problems caused by evolution of dust from the car loading station, from the cars during transportation and from the car unloading, coal stacking, storage, reclaiming and conveying operations at the conversion plant could be alleviated or eliminated.

Tests have shown that losses in transit can be substantial. Youngstown Sheet and Tube Company, for example, found that about 2,700 to 3,000 pounds of coal were lost as dust from 70 ton cars during the 200 mile transit from mine to point of use (8). This represents a weight loss of about 2 percent from the top layer of coal in the car. For a coal conversion plant receiving 18,000 to 20,000 TPD of coal in these cars, potential windage losses could be on the order of 120,000 to 135,000 tons year distributed along the right of way.

When the settling rate formula that was discussed earlier is applied to the railroad car problem, with H taken as 2 feet, V taken as 40 MPH and L of the car varying from 4 feet to 40 feet, the diameter of the largest particle that could be lost from the back of the car is about 470 micrometers, or about 32 Tyler mesh, and from the front about 148 micrometers, or about 100 Tyler mesh.

As a first approach to the problem, then, the statement may be made that if all particles in the loaded car were 32 mesh or larger, windage losses during transit would be reduced or eliminated. In handling and storage, with reference to the previous discussion on fugitive dust loss from storage piles, in a 20 MPH wind 32 mesh particles could be expected to reach the ground no further than about 45 feet from the peak of the pile.

Reduction of the dust content of the coal before it is loaded into cars has many advantages, among which may be listed:

- Fugitive dust losses at car loading are reduced
- Dust losses during transport are reduced
- Fugitive dust problems during unloading, storage, reclaim, and handling are reduced
- Need for dust suppression systems at car loading and at the conversion plant is reduced or eliminated
- Need for chemical binders on loaded cars in transit is reduced or eliminated

With reference to Figure 10-1, the crushed coal at the mine may be estimated to contain on the order of 1.4 to 6.5 percent of particles passing 32 mesh. Separation of these particles may be accomplished by wet or dry screening or by wet or dry elutriation. Agglomeration may be by compaction or granulation, with the object of agglomerating the fines into masses having a nominal diameter of 32 mesh or larger or a range of about 20 to 32 mesh (833 to 470 micrometers). The estimated load on the agglomeration step would be 14 to 65 TPH for a 1,000 TPH coal preparation facility.

For compacting, the particles are squeezed between two revolving cylinders, or rolls. Roll separating force, maintained hydraulically or by springs, may range up to about 400 tons. Capacity is a function of peripheral speed of the rolls, the spacing between rolls and the roll face width. A compactor with 30-inch face rolls of 6-foot diameter running at 120 RPM with an 800 micrometer spacing will compact on the order of about 18 TPH of fines, using about 30 horsepower. The compacted material is usually broken in a saw tooth crusher and screened to remove fines. Fines are recycled to the compactor. Assuming 15 percent of the compactor discharge will be fines, the maximum total feed to the rolls would be 75 TPH and a battery of 5 mills would be needed to handle the load.

Briquettes may be produced in equipment that is similar to the smooth roll compactor except that the rolls have depressions in their faces into which the dust is squeezed and formed into pillow or almond shapes. Commercial machines usually produce briquettes of one inch diameter or larger. Because the briquettes are large, capacity of the machines is greater, with respect to roll diameter and length, than the smooth roll compactors. For example, a briquetting machine manufactured by K. R. Komarek Inc. has rolls 36 inches in diameter and 12 inches wide, a roll separating force of 450 tons and a capacity of about 40 TPH, using about 75 horsepower. On this basis, two briquetters would suffice to process the fines from production of 1,000 TPH of coal.

Depending on such coal characteristics as the size consist, moisture content and susceptibility to plastic flow under pressure, the requirements for addition of moisture and a binder, and the amount of each, will differ for different coals in the compaction and briquetting processes. Addition of moisture leads in most cases to a requirement for a drying step which, in addition to removing moisture, also hardens the agglomerates. Binders may be needed, such as bentonite clay, lignosulfates that are byproducts of wood pulp manufacture, pregelatinized starches, causticized humic acids, fuel oil, waxes, asphalt,and pitch, and may be used alone or in combinations.

Coal fines may be granulated either on an inclined rotating disc or in a horizontal drum. Fine, free-flowing materials are fed to the disc or drum at a uniform rate and are sprayed with a binder that is usually a water solution, but can be a melt of wax or asphalt or similar material. Small pellets form as a result of the continuous rolling, tumbling action in the granulator and serve as nuclei for further pellet growth by layering of the fines and binder on the nuclei.

The inclined disc has an inherent classifying action that separates fines feed, nuclei, and product pellets into distinct zones. Therefore, only the desired size range of pellets is discharged and production of off-size pellets is minimized. Pellet size is controlled by such variables as pan inclination, pan speed and the location of the feed point and the binder sprays. Product pellet sizes usually range from 0.125 to 1.0 inch in diameter. For coal dust granulation, however, micro pellets in rather irregular shapes less than 0.125 inch (131 micrometers or 6 Tyler mesh) can be produced with proper choice of operating variables. Where small sizes are desired and the tolerable size range is wide, capacity of a disc increases. For example, it is estimated that a 20-foot diameter disc would produce only about 40 to 45 TPH of 0.5-inch pellets, requiring 125 horsepower, but would produce over 65 TPH of 6- to 20-mesh pellets with the same drive power. Therefore, one 20-foot disc appears to be adequate to granulate the dust from a coal preparation plant producing 1,000 TPH.

The drum granulator builds pellets in the same manner as the disc. Because the material flow through the nearly-horizontal cylinder is much less controllable than the flow across a disc, the size range of the pellet discharge is much wider with the drum. Firmness of the pellets may vary considerably, inversely with diameter, because of exposure of the pellet nuclei to varying amounts of binder and fines feed. There is almost always unpelletized dust in the drum discharge. This requires that the product be screened with subsequent recycle of fines. If a recycle ratio of 2:1 is assumed, the total throughput for a drum granulator would be 195 TPH in order to produce 65 TPH of granulated fines. To process this quantity of material, 2 drums would be needed, each 11 feet in diameter and 30 feet long and requiring about 75 horsepower to run at about 8 to 9 RPM. Bed depth in each drum would be about 3 to 3.5 feet. Retention time would be about 5 minutes.

Comparison of roll compaction, briquetting, disc granulation,and drum granulation leads to general conclusions as follows:

- Power requirements for the four types of agglomeration machines are approximately equal at 150 horsepower for production of 65 TPH of agglomerates

- Auxiliary equipment is required in compaction and drum granulation for separation and recycle of unagglomerated or created fines

- Presence of unagglomerated fines in the products of disc pelletizing or of briquetting is probably insignificant and separation probably would not be required

- Although meaningful capital and operating costs for the four types of processes were not available, it appears that the least costly, per ton of agglomerated fines produced, would be disc pelletizing, followed in order of increasing cost by drum granulation, compaction,and briquetting, with possibly little real difference between the last two.

- Disc pelletizing appears to be superior to other agglomeration methods from the standpoints of simplicity, ease of control, suitability of product, and costs

Coal fines may be pelletized on a disc by addition of water alone. Product size in this case is a direct function of product moisture content. For the micropellets desired for dust elimination, the optimum moisture content is on the order of 20 percent, with variation depending on the size consist and the coal type. Because of the low water solubility of the constituents of the coal, there is little residual tendency for the individual particles to adhere to each other when the moisture is removed by drying, with the result that the pellets tend to be fragile and to disintegrate readily.

Use of a binder improves the strength of pellets made by water addition (9):

- Bentonite is the most effective, but is expensive and increases the ash content of the coal

- Liquid lignosulfates, byproducts of wood pulp manufacture, are lower in cost than bentonite and do not increase the ash content of the coal. They are water soluble, however, and the upper layers of pellets in outside storage or transport will deteriorate. They are usually applied at a rate of 1.0 percent by weight of dry solids

- Pregelatinized corn flour starch is highly effective as a binder and, when dried, is not affected by rainfall on the storage pile or in transit, but is more expensive than the lignosulfates

- A mixture of equal parts of the starch and lignosulfates applied at a rate of 0.5 to 1.0 percent by weight of solids has been shown to be effective

- A mixture of equal parts by weight of dry solids of lignosulfates and bentonite, applied at a rate of 0.5 to 1.0 percent by weight of dry solids, resulted in excellent pelletization and little increase in the coal ash content

- A binder made by reacting humic acids with calcium hydroxide has been shown to be as effective as bentonite in the pelletization process and in final pellet strength and costs much less than bentonite

- In general, increasing the proportion of binder yields stronger, but larger, pellets

Use of water and binders is effective in pellet production but requires a drying step to reduce the pellet moisture content to about 2 to 3 percent and to increase pellet strength. High drying rates have been shown (9) to decrease final pellet strength. Although pellets of 0.5 inch diameter and larger may be dried in conveyor type thermal dryers, supported on wire mesh for through circulation of hot air, the micropellets that are under consideration here would require, for example, fluid bed drying for least attrition.

Drying may be eliminated if the coal fines contain little water as fed to the pelletizer and if a nonaqueous binder is used. Tars, pitches,and waxes are examples of such binders that may be applied to the pelletizer through sprays as melts, singly or in combination, or in mixtures with fuel oil. The product pellets would need only cooling to become durable enough for transport, and would be waterproof. Elimination of the drying step should compensate, at least in part, for the increased cost of the binding agents.

Attention has been given so far only to the fines that are created during the coal crushing operations. Where coal fines are recovered as a slurry from flotation or from settling ponds and are filtered for water separation, the moisture content of the damp mass of solids will be on the order of 25 to 30 percent (10). With admixture of a binder, and of dry fines to adjust the liquid content to the proper level for the desired degree of pelletization, these solids may be pelletized and dried to increase the overall yield of the mining operation.

Where both dry and wet fines are to be pelletized, an economic study would be required to determine the best process route, given the following choices:

- Two trains, one handling dry fines with a melt type binder and no dryer, the other handling wet fines with a solution type binder and a drying step
- Mix the wet and dry fines and feed to discs for water and binder pelletizing followed by drying
- Partial drying of the wet fines and mixing with the dry fines so that the total mixture has a moisture content of about 3 percent, then pelletizing with a melt type binder and no dryer.

Consideration may also be given to use of one or more of the chemical binders described previously under the heading "Dust Control." Experimentation would be needed to optimize the scheme, with cost as the criterion.

Collection and Disposal of Coal Dust

Coal is reclaimed from storage by several means, all of which are potential dust generators:

- Through openings under the piles leading to feeders that discharge to conveyors
- By bucket wheel reclaimer, in which toothed buckets are attached to the rim of a large wheel that rotates on a horizontal axis. The wheel assembly is mobile so that it may be advanced into the pile or moved parallel to it. The buckets scoop up the coal and dump it onto a conveyor
- By trencher, similar in action to the bucket wheel except the toothed buckets are mounted on a traveling chain

Dust is usually controlled at the reclaiming station by water sprays that may contain a surfactant, as described in preceding paragraphs. Coal is transported by conveyors to the conversion process feed preparation area. Covering of the conveyors will normally prevent fugitive dust loss and, if the coal is wetted during reclaiming, dust generation at transfer points will be minimized.

At the primary crushing station the coal passes across a scalping screen, or grizzly, where the small particles of coal are separated from the larger lumps. The primary crusher reduces the large lumps of coal to a size consist that, depending on the conversion process, may be fed to the reactor without further processing or may be fed to a pulverizer. The discharge of the crusher is combined with the finer fraction that passed through the grizzly and the whole stream is (usually) transported to live storage silos. In the primary crushing area the discharge of the feed conveyor, the grizzly, the crusher and the conveyor feed point for the intermediate size coal product are all enclosed and vented to a dust collection system. Pickup points for evolved dust are strategically located in the crusher area. Air velocity at the openings of the pickup nozzles varies from a minimum of about 200 up to about 400 feet per minute, depending on, for example, the coal type, range of moisture content and the particle size distribution at the points in the system. Air velocity in the ductwork connecting the pickup points to the baghouse is normally 2,500 to 3,000 feet per minute to prevent settling of the dust particles.

A similar system is used to control dust in subsequent transport and grinding/pulverizing operations

Baghouse collectors are used almost universally to separate coal dust from the transporting air, since coal dust particle sizes range upward from about 2 micrometers and for this duty the collection efficiency is rarely less than 99 percent. Although the capital cost of bag collectors is high and they require the most space for installation of any of the various types of dust collection equipment, they require much less energy than the other types to achieve their high-efficiency recovery and require no water for operation. In the dust collection system the bag collectors may be preceded by cyclone collectors which will remove most of the dust particles of 10 or more micrometers diameter and will thus relieve the bag collector of part of the load.

Storage silos are either vented separately through small, individual bag collectors or air filters or are vented together through a common bag collector. The individual collectors discharge their collected dust into their respective silos. The common dust collector may discharge to the orginal silos or the collected dust may be combined with the dust from the transport and milling operations. In the latter system, all collected dust may be conveyed, usually pneumatically, to a point in the conversion plant coal preparation system where it may be conveniently combined with the main stream of the coal feed. As an example, the pneumatic conveyor may discharge into the top of a surge bin, from which the dust may be fed as needed, and in which a fabric filter separates dust from the transporting air before the air is discharged to atmosphere.

Alternately, the dust may be slurried with water or oil and fed to the conversion process reactor, as in the Texaco gasification or the SRC liquefaction processes, respectively.

In the Lurgi Dry Ash process the coal feed can contain no fines. In this case the process feed is screened to remove fines and fines and collected dust are fed as fuel to the incinerator/boiler that is part of the process scheme.

Control of Other Process Dusts

Ash and slag from the coal conversion processes are quenched as they are discharged from the reactors. Much of the free water is removed during passage of the ash slurry through hydrocyclones and sieve bends until the combined ash stream from conversion and from steam generation contains on the order of about 20 to 25 percent water. In this form the solids stream is easily transported by conveyor belt or in bulk carriers and is not dusty while the surface layer is wet. Dust control while the material is within the process battery limits may therefore be limited to wetting devices such as sprays or high pressure hose streams on any piles of slag or ash awaiting transfer to permanent disposal areas.

Limestone for flue gas desulfurization may be received as lumps for eventual in-plant grinding or as powder for direct feeding to the scrubbers. Dust control methods in handling of lump limestone from car unloading through storage, reclaim and primary crushing may be similar to those in coal handling: wetting the particles during unloading with water containing sufficient surface active agent to provide a carryover effect and supplementing the wetting at unloading with wetting at transfer points and during reclaim. Following reclaim, control of dust evolved during crushing, screening, fine grinding, silo storage, reclaim,and feeding to the flue gas scrubber system can be accomplished dry, by bag collector, or wet, by wet cyclone,or other wet collectors. If wet collectors are used, liquor from the flue gas scrubber may be the collecting medium, for example, with the collected dust being discharged to the scrubber system as part of the feed.

Where ground limestone is received in carload lots, the cars may be unloaded pneumatically or by dumping into under-track hoppers. With the pneumatic system, cars of special design ("Airslide," for example) discharge into closed systems that do not permit dust to escape. The ground limestone is finally discharged into closed silo storage, vented through fabric filters. Dust control is a greater problem when cars dump into under-track hoppers and the limestone is to be transported, stored,and fed as a powder: since wetting is not permitted, dust must be collected by air streams and then separated from the air in a bag collector. The volume of air to be handled can be minimized by reducing to a practicable minimum the open area between the car and the hopper and sealing the hopper bottom to the powder transport mechanism, which can be a screw conveyor or a pneumatic conveyor.

As an alternate in powdered limestone handling, the cars could dump into receiving tanks to become slurries containing on the order of 30 to 50 percent solids. The slurrying medium could be either makeup water for the scrubbing system or scrubbing system liquor. In either case, sprays of the liquid could be used to control dust evolution.

Spent catalysts from conversion processes are changed periodically because of deactivation through poisoning. Quantities are small in comparison with other solids streams, but dusts evolved in catalyst handling are hazardous and/or toxic and require control. Examples of the catalysts are:

- Cobalt-molybdenum, catalyst for naphtha hydrogenation
- Methanation catalyst for SNG production
- Shift conversion catalyst for hydrogen production
- Claus catalyst for sulfur production
- Copper-based catalyst for flue gas denitrification
- Desulfurization and denitrogenation catalysts

Catalysts are unloaded from their containing vessels in the conversion process sequence by dumping or by hand raking and are received in covered containers. Dust control is accomplished by small, individual mobile units consisting of fans, bag filters (or other fabric medium),and appropriate flexible ductwork. Collected dust is placed in the covered catalyst containers. The containers are trucked either to the solid waste disposal area, to be mixed with the ash/slag stream at a convenient point, or to the pug mixer in which quenched ash/slag, ash,and evaporator bottoms are combined for transport to disposal. In either case, the catalysts are dispersed in the damp solid wastes and dusting beyond the dispersal point is not a problem.

COSTS OF DUST CONTROL

Dust Suppression by Water Sprays

Comparison of water spraying systems at the coal handling facilities that were visited during this project, and conversations with suppliers of spraying systems, lead to the conclusions that there is little uniformity in design and installation of the the spraying systems, that despite the "custom design" of any one installation there is little or no attempt at scoping or defining the individual dust problems by either the plant operator or the spraying systems designer and that, finally, dust suppression sprays are installed by rule of thumb with sufficient spare capacity so that if suppression is not achieved at the design spraying rate, then the rate may be doubled or trebled to achieve suppression. Capital and operating costs for spraying systems are difficult to evaluate because of these many variables and empirical factors in system design and operation.

A study leading to the determination of spraying systems costs was carried out in 1974 by the Bureau of Mines for three sizes of hypothetical hard rock crushing plants (6). While the conditions in rock crushing operations may not exactly parallel those in coal handling in conversion plants, there is sufficient similarity to warrent cautious comparison. For example, the assumption in the Bureau of Mines study that 1.5 gallons of surfactant solution, with a surfactant to water ratio of 1:1,000, would be required per ton of rock processed is in the same range of solution usage (1.5 to 2.0 gallons per ton) that is stated as "normal" for coal handling.

In the Bureau of Mines study the rock crushing plant operated 1,920 hours per year at 1,000 tons per hour. The surfactant solution was sprayed at 1.5 gallons per ton. The spraying system therefore had an operating output of1,500 gallons per hour of solution. In contrast, a coal conversion plant producing 250 billion Btu per day of gaseous or liquid fuels will require on the order of 20,000 tons per day of coal. If it is assumed that the coal feed will be unloaded during 2 shifts per day for 6 days per week during 48 weeks per year, then the coal unloading rate becomes 1,435 tons per hour. If it is further assumed that the coal will require 2 gallons of surfactant solution per ton of coal, then the operating output of the spraying system will be 2,870gallons per hour, or 1.9 times the capacity of the system in the Bureau of Mines study.

The installed cost of the Bureau of Mines spraying system, including winterization, is shown as $61,676. When the 0.6 exponent is used to scale up the capital cost for the increase in the capacity and 17 percent is added for escalation of costs from mid 1974 to the end of 1977, the capital cost for the coal spraying system becomes $106,100. In both cases it is assumed that these are turnkey installations and that electric and water utilities are in place to supply the systems.

Operating costs are calculated on the basis of operating 2 shifts per day, 6 days per week, 48 weeks per year, or 4,608 hours per year. Because the system is controlled automatically in conjunction with operation of the coal reception station, operating labor is minimal and is estimated at 8 hours per week or 384 hours per year. Maintenance labor in the Bureau of Mines study was estimated at 8 hours per week for 40 hours per week system running time. For the coal system, with 96 hours per week operating time, the maintenace labor is estimated to be 16 hours per week or 768 hours per year. Operating costs for the coal spraying system calculated from these bases are shown in TABLE 10-4.

TABLE 10-4. CAPITAL AND OPERATING COSTS FOR WET DUST SUPPRESSION

Basis:	Coal rate	1,435 TPH
	Spray rate	2 Gal/T
	Operation: 2 Shifts/Day, 6 Days/Wk, 48 Wks/Yr	
	Depreciation: 20 Yrs., Straight Line	
	Fixed Capital	$106,100
	Working Capital (60 Days Operating Cost)	19,400
	Total Investment	$125,500

Operating Costs per Year

Power 10 hp x 4608 hrs. x 1.8¢/Kwh	$ 622
Operating Labor 384 hrs. x $10	3,840
Maintenance Labor 768 hrs. x $10	7,680
Maintenance Materials	7,000
Wetting Agent 0.002 gal/ton x 6,600,000 x $3.05/gal	40,260
Overhead 35% of labor	4,032
Indirects 40% of maintenance and supplies	21,976
Taxes and Insurance 2% of fixed capital	2,122
Depreciation 5% of fixed capital	5,305
TOTAL ANNUAL OPERATING COST	$92.837
OPERATING COST PER TON OF COAL	$0.014

Dust Control by Chemical Binders

Application of chemical binders to the outside surfaces of dead storage piles has been shown to reduce the problem of fugitive dust (7). The chemicals in water solution are sprayed on the piles and form stable, inert, non-toxic and non-flammable films. Spraying is accomplished with conventional equipment, such as that used for spraying orchards, or with custom built spray trucks with large capacity tanks, high pressure pumps and special nozzles.

A stockpile containing about 15 days supply of coal for a 250 billion Btu per day conversion plant, or 300,000 tons, may consist of 4 piles, each 45 feet high and about 1600 feet long. Total pile area to be coated in this configuration is about 825,000 square feet.

Compound SP, made by Johnson-March Corporation for coating piles, is a synthetic organic polymer in a water base. It is used as received, without dilution. Usual application rate is one gallon per 100 square feet of surface area. Effective protection is provided for about one year. Various grades of the material have been developed to cope with such variables as differences in types of materials to be coated, particle size, surface wettability,and climatic conditions. Cost of Compound SP is about $2.50 per gallon, and therefore the cost of the 8,250 gallons of the compound to coat the total pile surface of the stockpile is about $20,600.

Coherex, manufactured by Golden Bear Oil Company, is an emulsion of petroleum resins, wetting agents,and sequestering agents in a water base. The usual dilution for use is one part Coherex to 4 parts water. The rate of application varies with the characteristics of the material being sprayed, but an average rate of

2 gallons of diluted Coherex per square yard, or 22 gallons per 100 square feet, may be considered. On sandy materials the penetration is about 2 inches at this rate of application. Cost of the Coherex concentrate is about $0.26 per gallon. About 4.5 gallons of concentrate are required to treat 100 square feet; therefore, the cost of 37,125 gallons of Coherex for one application to the total surface of the stockpile is about $9,700. It is probable that coating would be needed twice a year, bringing the annual cost of material to about $19,400.

A special tank truck carrying high pressure pumps and spray nozzles may cost about $30,000, fully rigged. Amortized over 6 years yields an annual cost of $5,000.

Spray rate for the undiluted Compound SP is estimated at 10 gallons per minute, at high pressure to atomize the spray, requiring about 14 hours to coat the pile. Spray rate for diluted Coherex is estimated at 40 gallons per minute at a pressure of 40 to 60 psig, requiring about 75 to 80 hours to coat the pile. If it is assumed that spraying will be done during 6 of the 8 hours of a shift, then 2.5 days once a year are needed for Compound SP and 12.5 to 13.5 days twice a year are needed for Coherex. Operating costs for the two coating systems may be compared:

	Compound SP	Coherex
Amortization of spray truck	$ 5,000	$ 5,000
Labor: 3 operators at $10/hr 2.5 days/yr	600	
26 days/yr		6,240
Maintenance labor	1,500	1,500
Fuel and other supplies	400	2,000
Coating compound	20,600	19,400
Taxes and insurance	1,000	1,500
Overhead 35% of labor	735	2,709
Indirects 40% of maintenance and supplies	9,000	9,160
TOTAL ANNUAL OPERATING COST	$38,835	$47,509
OPERATING COST PER SQUARE FOOT COATED	$0.047	$0.058

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SOLID WASTE DISPOSAL AND MANAGEMENT

Problem Definition

Coal conversion processes generate solid wastes with widely divergent characteristics:

- Ash from gasifier reactors that operate with a maximum temperature below the ash fusion point is composed of loose agglomerates of small particles that break readily into discrete particles. Average bulk density of the dry solids is 40 to 45 pounds per cubic foot. Dry particles easily become airborne. Gasifier ash is quenched with water in all conversion processes. The resulting slurry, if it is allowed to settle undisturbed in ponds, will eventually compact to about 25 pounds of dry solids per cubic foot (1,p.IV-18) with an equivalent water content of about 65 percent. When subjected to classification, thickening and filtration, however, the water content can be reduced to 15 to 20 percent

- Ash from gasifier reactors that operate with a maximum temperature above the ash fusion temperature is discharged from the reactor as a melt. During subsequent water quenching the melt shatters into discrete particles resembling sand. Average bulk density of the dry quenched solids is about 100 pounds per cubic foot. Dry particles become airborne only with difficulty. Solids in the slurry from the quenching step may be separated from the liquid by classification, thickening, and filtration to yield a final stream containing only about 10 to 15 percent water

- Ash discharged from utility boilers or incinerator/boilers may be a combination of bottom ash and flyash or may be bottom ash alone, depending on the boiler configuration and method of operation. In these cases the ash characteristics are similar to those from low temperature gasifiers. With the increasing attention being given to fluidized bed combustion, in which limestone is mixed with the coal fuel to act as an absorbent for sulfur dioxide, the ash may be a mixture of "normal" ash, calcined limestone, calcium sulfite and calcium sulfate. Quantity will depend on the utility demand of the conversion process plant and the proportion of the total heat requirement that is furnished by incineration of organic residues from the conversion process. Physical characteristics will, in general, be similar to "normal" boiler ash. The total ash discharge may be sluiced to the ash handling system or it may be quenched only enough to reduce the temperature to the range than can be tolerated by conventional transport equipment

- Flue gas desulfurization (FGD) sludge varies greatly in composition depending on the coal composition, the degree of fly ash removal prior to scrubbing, scrubber operating characteristics, and the type and amount of additives. Sludge from a lime/limestone wet scrubber is mainly calcium sulfite, calcium sulfate, calcium carbonate and ash. The sludge may be dewatered to an average solids content of about 50 percent and an average bulk density of around 70 pounds per cubic foot. In settling ponds there is less dewatering and the fluid settled sludge may contain considerably less than 50 percent solids. Both dewatered and settled sludge have low bearing and compressive strength and are thixotropic

- Organic sludges from biological oxidation processes are susceptible to bacterial degradation when they are exposed to air, or possibly to anaerobic fermentation if they are covered. Although the quantity of organic sludge that is generated in coal conversion plants is small in comparison to the quantity of ash and FGD sludge, odors originating in the organic sludge could become a nuisance and, possibly, a legal problem. In the treatment schemes that have been examined in this study, the preferred method of disposal of organic sludges is incineration. This has been shown in Section 8 in the integrated schemes for Lurgi gasification and SRC liquefaction. Bi-Gas and similar high temperature gasification processes do not require biological oxidation in the water treatment sequences and therefore do not produce organic sludges.

- Inorganic salts accumulate in boiler and cooling tower blowdowns and originate in water treatment processes. As proposed in the integrated schemes for water treatment in

Section 8 of this study, the inorganic salts may be concentrated by evaporation. The constituents of the concentrate will vary widely, depending on raw water analysis and methods of treatment of the raw and recycling water. The concentration of the salts in the evaporator bottoms will vary depending on the limitation of concentration due to scaling of heat exchange surfaces and the type of evaporator. For example, the salts may be evaporated to near dryness in an oil suspension, then may be separated from the oil by centrifuging. The quantity of inorganic salts, on a dry basis, is small in comparison to the quantity of ash. The possibility exists that the inorganic salts may react with components of the ash or FGD sludge to yield soluble compounds. In any case, the evaporator bottoms will probably contain both soluble and insoluble compounds.

- Spent catalysts are discharged periodically in quantities that are small in comparison to the ash stream. They contain metals or metallic compounds that may be harmful to the environment, such as cobalt, molybdenum, copper, and zinc and may during their active life pick up other potentially harmful metallic compounds. The spent catalysts are discharged dry from their containing vessels in the conversion process sequence.

The Pullman Kellogg study of control of gaseous emissions (Section 9 of this report) did not consider flue gas desulfurization by lime/limestone scrubbing, preferring instead to take advantage of the rich hydrogen sulfide stream that is fed to Claus sulfur recovery to supply the reducing gas to the citrate process sulfur recovery unit on the incinerater/boiler. By so combining the two sulfur recovery systems the total amount of sulfur for sale is increased, the capital cost for emission control is reduced and overall operating costs for control of sulfur compound emission is reduced.

If, however, FGD by lime/limestone scrubbing were the only means of controlling sulfur compound emissions from coal conversion plants, production of FGD sludge may be estimated from the following information, which is based on power plant operation (1, p.IV-32ff.):

- 95 percent of the sulfur in coal fed to boilers becomes sulfur dioxide. The rest of the sulfur remains with the ash

- Density of dewatered sludge is 57 to 85 pounds per cubic foot and averages about 71 pounds per cubic foot

Since about 93 to 98 percent of the sulfur compounds in coal fed to conversion processes are converted to hydrogen sulfide which, if not recovered, must be oxidized to sulfur dioxide and removed by FGD scrubbing, for the purposes of determining the amount of FGD sludge and evaluating its effect on the overall solid waste management scheme the conversion processes can be considered to be roughly equivalent to boiler operation. Accordingly, the variation in ash and FGD sludge production with changes in coal composition may be evaluated as in TABLE 10-5.

If sulfur is recovered via the Claus process from streams rich in hydrogen sulfide, all other sulfur-bearing streams, including Claus tail gases, are oxidized to sulfur dioxide, and FGD scrubs the gases, the dry solids in FGD sludge become:

	FGD Solids	Sludge/Ash Ratio
Gasification, high sulfur coal	1,119 TPD	0.60
Gasification, low sulfur coal	244	0.16
Liquefaction	1,288	0.62

Maximum sulfur recovery in the conversion processes, including

TABLE 10-5. ASH AND FGD SLUDGE PRODUCTION WITHOUT SULFUR RECOVERY

Basis: Production of 250 billion Btu/day of gaseous or liquid fuels

Feed sulfur to gasification FGD scrubbing:

High sulfur coal	98.6%
Low sulfur coal	95.0%

Feed sulfur to liquefaction FGD scrubbing:

High sulfur coal	93.4%

		Gasification		
		High Sulfur Coal	Low Sulfur Coal	Liquefaction
Coal feed:	Ash, %	11.30	7.72	11.80
	Sulfur, %	4.42	0.66	3.70
	Btu/lb	13,190	11,290	12,125
	TPD	14,778	17,265	17,680
Ash production, TPD		1,851	1,553	2,070
Sulfur to FGD, TPD		644	114	611
FGD Sludge, TPD dry solids*		4,025	713	3,819
Ratio, Sludge/Ash		2.17	0.46	1.84

* Typical composition:	$CaSO_3 \cdot 0.5H_2O$	58%
	$CaSO_4 \cdot 2H_2O$	9%
	$CaCO_3$	33%

Sulfur Content = 16%

sulfur recovery from stack gas scrubbing, will yield no FGD sludge.

These comparisons illustrate the extreme variations that may be encountered in coal conversion plants, depending on the economics of sulfur recovery vs. sulfur disposal.

With reference to Figure 8-54, the water balance for Lurgi gasification of low sulfur coal, the solids-bearing streams are shown to be:

	Total	Solids	Water	% Solids
Evaporator bottoms, TPD	153.6	18.0	135.6	11.6
Boiler ash, TPD	264.0	264.0	0.0	100.0
Superheater ash, TPD	18.0	18.0	0.0	100.0
Dewatered gasification ash, TPD	1,495.3	1,271.0	224.3	85.0
Total to disposal, TPD*	1,930.9	1,571.0	359.9	81.4

*Biological oxidation sludge not included

If there is no sulfur recovery, the FGD sludge at 50 percent solids added to the above total yields a total mixture of 3,414.6 TPD, containing 2,284 TPD of solids, or about 68 percent.

With high sulfur coal in Lurgi gasification, assuming no change in evaporator bottoms, boiler ash and superheater ash, the total solids to disposal becomes 2,351.4 TPD containing 1,869.0 TPD of solids, or 80.2 percent. Addition of FGD sludge yields a total mixture of 10,401.4 TPD containing 5,912 TPD of solids or 56.8 percent.

With high sulfur coal feed to liquefaction, maximum sulfur recovery and no FGD, reference to Figure 8-58 shows that the total to disposal is 2831.7 TPD, containing 2,123.8 TPD solids, or 75 percent. If maximum FGD sludge is added, the total stream to disposal is 10,469.7 TPD, containing 5,942.8 TPD of solids, or 56.8 percent.

As previously noted, the dry bulk density of the ash discharged from gasifiers operating below the fusion temperature of the ash is about 40 to 45 pounds per cubic foot. The density may be increased to about 60 to 65 pounds per cubic foot by compacting the disposal pile. In contrast, the ash from gasifiers operating above the fusion temperature of the ash, and the ash from boilers and incinerators, will have a dry uncompacted bulk density of about 100 pounds per cubic foot, and a compacted bulk density of about 110 pounds per cubic foot. A bulk density of 130 pounds per cubic foot may be assumed for the solids in the evaporator bottoms. The total volume of uncompacted solids discharged from coal conversion processes without FGD may then be estimated as shown in TABLE 10-6. Uncompacted densities were used so that structural considerations of the disposal areas could be simplified.

From TABLE 10-6, the volume of the solid wastes may range from 31,400 to 75,600 cubic feet per day, depending on the coal type and the conversion process operating characteristics. Considering operation at 330 days per year yields an estimated solids volume of 10.4 to 25 million cubic feet per year.

Solids Transport

The solids mixture as it is produced in the conversion plant, without FGD sludge, is a damp mass that may be transported by conveyor belt or by truck to the disposal area. Choice of transport method would be based on economics, considering the distance between plant and disposal area, the terrain and

TABLE 10-6. VOLUME OF SOLIDS FROM COAL CONVERSION WITHOUT FGD

	Low Temp.	High Temp.	Liquefaction
Ash from gasifier:			
TPD	1,271-1,569	1,218-1,452	590
Density, PCF*	45	100	100
Volume, MCFD*	56.5-69.7	24.4-29.0	11.8
Ash from boilers:			
TPD	282	335-399	1,480
Density, PCF	100	100	100
Volume, MCFD	5.6	6.7-8.0	29.6
Evaporator bottoms:			
TPD	18	18	54
Density, PCF	130	130	130
Volume, MCFD	0.3	0.3	0.8
Total dry solids, TPD	1,571-1,869	1,571-1,869	2,124
Total solids volume, MCFD	62.4-75.6	31.4-37.3	42.2
Moisture in total solids, %	19-20	30	25
Bulk density of wet solids, PCF	62	143	134

*PCF = pounds per cubic foot of compacted solids
MCFD = thousands of cubic feet per day

climatological conditions. Trucking stabilized FGD sludge at 60 percent solids is evaluated in a TVA study (2). In the study capital costs were developed for mid-1979 for trucks and earthmoving equipment to haul about 2,400 TPD of sludge 1 mile to disposal for the IU Conversion Systems (IUCS) Poz-O-Tec process:

Hauling Distance, mi	Capital Cost for Transport
1	$581,000
3	617,000
5	641,000
10	700,000

Because of the tendency of wet, stabilized FGD sludge to dewater during truck transport and cause spills on the roads to the disposal site, hauling distance must be limited to 2 to 3 miles, according to an evaluation by Columbus and Southern Ohio Electric (3). In C&SOE's operating system, freshly-mixed stabilized FGD sludge is carried by conveyor belt to a radial stacker which forms a surge, or curing, pile from which earthmoving equipment pushes the partly cured Poz-O-Tec stabilized material to final disposal, where the sludge is laid down in layers about 2 feet thick and then compacted to a dry solids density of about 65 pounds per cubic foot. Projected final height of the pile is 100 feet.

These findings on transport of stabilized FGD sludge may be applied to transport of ash and the other solids from the conversion plant to the disposal site. The problems in ash transportation appear to be less severe than with stabilized FGD sludge:

- Solids content of the mixed solids, as shown in TABLE 10-6, varies from 70 to 81 percent

- The ash mixture is not thixotropic, as untreated, freshly-mixed or partly cured FGD sludge is, and therefore leakage from trucks during road hauling should be greatly reduced. Special construction, such as gasketed tail gates on dump bodies, could eliminate leakage.

- Transport by conveyor belt and stacker has been demonstrated and is in operation for freshly-mixed stabilized FGD sludge. The transport problem is greatly reduced for the ash mixture because of its higher solids content and lack of thixotropic tendencies. Further, transport by conveyor of wet ores of various particle sizes is common commercial practice.

Ash and mixed solids may be transported by aerial tramway (5). Tramcar capacities vary from 10 to 90 cubic yards and are able to travel at rates up to 1750 feet per minute and seldom under 1,000 feet per minute (4). Since spans of 1500 feet or more between towers are being operated, the system may be considered as a viable means of transport and practical from the standpoint of being able to span the disposal area at any desired height. Towers 350 feet tall are in operation. Considering the solids from low temperature gasification, as shown in TABLE 10-6, loaded into 90 cubic yard tramcars, each car would carry 75 tons of wet solids or 60 tons of dry solids and 31 carloads per day would satisfy the plant. If the number of cars is conservatively estimated at 35, if operating time for the tramway is 7 hours per day and if the traverse rate is 1,000 feet per minute, then the disposal area could be located over 2.25 miles from the conversion plant. Variations on this scheme would require study to determine the most economical combination of the following:

- Operate 24 hours per day with smaller tramcars to reduce cable loading and capital cost. Also reduces stockpiling of mixed solids awaiting transport to a minimum surge pile

- Possible advantages of increasing tramcar speed, or the number of cars in transit in order to increase the practical transport distance

- Need for transfer points, and the required equipment, for long transport distance

- Hazards and nuisances involved in overhead transport of wet solids

Economic evaluation of belt conveyors as a means of transport must be carefully done, considering the following:

- Cost per ton mile tends to remain constant, no matter how far the conveyor is extended

- Distribution at the discharge end of the belt may be by truck, scraper or stacker units. Bins must be provided if trucks or scrapers are used, but are not needed with stacker units

- Belt conveyors may be combined with tramways for negotiation of difficult terrain

- In the belt conveyor layout provision should be incorporated for reverse return where the belt, after passing the discharge point, is twisted 180 degrees so that the wet side of the belt does not come in contact with the return idler rollers. Wear on the return idlers is reduced, buildup of sticky material on the return idlers, with consequent problems with maintaining belt alinement, is avoided and droppings of waste on the beltway are reduced

Wheeled vehicles may be used to transport and disperse waste. Rear dump trucks, side dump trucks, and scraper loaders have been used for coal refuse disposal (4) and would be suitable for conversion plant waste transport and disposal. These vehicles can spread the waste in a thin layer in the disposal area and can compact it by driving the vehicles over the area while the loads are being discharged. Vehicles have great flexibility in operation, the means to adapt the discharge pattern to changing contours in the disposal area and to develop area stability, together with their intrinsic flexibility in capacity that is characteristic of multiple vehicular transport. In addition, the cost per ton mile for wheeled vehicles tends to decrease as the haulage distance increases. Economic evaluation of vehicular transportation and distribution of conversion plant wastes must consider:

- Special vehicle construction that may be required by the characteristics of the material being handled

- Special construction that may be required so the vehicles can traverse the disposal area

- Environmental regulations that may limit the effectiveness of vehicles, such as restrictions on roads travelled, restrictions on spillage or droppings on roads, and restrictions on creation of hazards or nuisances from dust and noise

Management of the Solid Waste Disposal Area

In the discussion that follows the wastes to be considered are ash/slag and evaporator bottoms. As previously pointed out, the quantities of such miscellaneous solids as spent catalysts are small, or they are discharged only intermittently, and therefore would have little effect on the total waste quantity or on the physical or chemical characteristics of the waste.

Flue gas desulfurization (FGD) sludge is considered only incidentally in this discussion because the decision to include an FGD system in a coal conversion plant is site-specific and economically oriented. Information and data on means of disposal of FGD sludge have been gathered, however, and will be applied to solid waste management wherever such inclusion may be helpful in defining, or advancing solutions for, waste disposal problems.

Elimination of organic wastes from the solid waste stream avoids nuisance problems of odors caused by decomposition. It is assumed in this study that the organic wastes will be incinerated and that any residue will be inorganic and combined with the ash stream.

The following is quoted from reference 1, p. V-6:

> "The effects of sludge and ash production will include increased demand for discharge sites, disruption of natural wildlife and vegetation and possible contamination of local water resources. When solid wastes are to be transported for disposal to an offsite landfill by trucks, secondary transportation impacts may occur, including increased use of fuel and equipment, increased air pollutant emissions and increased traffic, dust, and noise in the area traversed. In developed areas, land will be converted from use in agriculture, housing, and industrial development to uses serving coal production and utilization. Land use will be permanently altered or disrupted when permanent facilities for coal transport and use are developed. Other areas used in mining and solid waste disposal may be reclaimed, eventually, but the time required and the degree of success of reclamation efforts are not known at this time. It is unlikely that natural topographical features and an exact replica of ecological systems will result from any reclamation program."

The quotation serves as a guideline for the discussion of management of solid waste disposal areas, together with the applicable environmental standards, both present and proposed.

A basic premise of the present study is that the control technology applied to coal conversion operations should be so chosen that gaseous emissions may be reduced to a level at or below the most stringent environmental standards that are now in force and that liquid effluents from the conversion processes should be treated for recycling within the process and should not be released to receiving waters. This premise has been presented, and the means to achieve it have been described in detail, in Section 8 of this report for liquids and Section 9 for gases.

Increasing the Solids Content of the Solids Waste Stream--

In the schemes proposed for treatment of liquids in coal conversion, commercial control technology was applied in various combinations to permit recycling the water. A residual amount of water remained, bearing inorganic salts, after evaporation to a concentration limited by the operating characteristics of the evaporator. As discussed previously, this evaporator bottoms stream was combined with quenched ash/slag from the gasifier reactor and the dry ash from the incinerator/boiler and the process steam superheater to yield a mixture contaioning 70 to 80 percent solids.

Although the damp ash mixture can be handled by conventional means, the solids content may be increased by use of an evaporating system marketed by Dehydro-Tech Corp., East Hanover, N. J., in which the water and solids are suspended in a circulating bath of oil (5). In this process, oil of the correct volatility, viscosity and surface tension is added to the aqueous waste and the mixture is fed to a multiple-effect falling-film evaporator. The mixture passes through the evaporator, leaving the solids suspended in the dry oil as a fluid slurry. The oil is then centrifuged off to be used again, and the dry solids are left. Water is recovered and recycled to the coal process, the oil is recycled, and the water-soluble solids emerge in a dry powder form.

The oil maintains fluidity throughout the system, improves heat transfer in the evaporator, and prevents scaling and fouling as the salts become completely dry. This patented technique is efficient for the removal of water without problems of thickening, scaling, and fouling.

It is reported that oil-fluidization plants have turned out to cost approximately one-third less than equivalent spray-drying or incineration plants, due mainly to economics of multiple-effect heat utilization. Evaporation is energy-intensive but the use of a strong brine as feed, along with multiple-effect heat cost savings, bring it within the limits of reasonable economy, given that clean-up of the aqueous waste is required.

Makeup oil for the process may be available as one of the products or byproducts of the conversion processes. Choice is governed by such oil properties as gravity, viscosity,and vapor pressure (boiling range). There is no theoretical reason against using the phenols fraction as the evaporator oil (5). This could be a small diversion from saleable products to be used as make-up oil, or it could be once-through on the way to incineration.

The oil, of whatever type, withheld from sales would be only make-up. Most of the oil would be recycled into the evaporation cycle, and the oil purge, if viscous impurities collect, would be recycled into the coal conversion process. The vapor pressure of the oil is such that about one pound of oil is vaporized with every pound of water. This can be diminished or augmented as desired by selecting oil fractions of appropriate boiling range. The cooled oil can then solvent-extract organics out of the water, which is then treated further or recycled to the coal process as needed.

For evaporation equipment, cost has been explored (1976) as affected by feed concentration and feed rate. The economics of scaleup in feed rate are apparent from the figures summarized in TABLE 10-7. It is also apparent that pronounced economies result from designing for the higher feed concentrations. These effects should be analyzed in an economic study to demonstrate the interaction of oil-fluidized evaporation with the overall water treatment system and the subsequent solids handling operations.

The advantage of oil-fluidized evaporation is that, with such an operation to remove water, the solids content of the mixture from low temperature gasification may be increased to about 87.5 percent, that from high temperature gasification to about 84.3 percent and that from liquefaction to about 89 percent. This reduction in water content further reduces any solids handling problems.

The decision to install oil-fluidized evaporation instead of conventional evaporation, or for there to be any evaporation step at all, depends on the following:

- Whether or not FGD sludge will be part of the solid waste stream

- Economics of transporting very wet (50 to 60 percent solids), damp (20 to 30 percent solids) or semi-dry (10 to 15 percent solids) wastes to disposal

TABLE 10-7. CAPITAL COSTS OF OIL-FLUIDIZED EVAPORATION (5)

	Demonstration Unit		Commercial Unit	
Feed Concentration, Wt Salts	20	1.32	20	1.32
Feed Rate, Pounds/Hour	8,000	127,868	40,000	631,000
Number of Effects	3	6	4	6
100 PSIG Steam, Pounds/Hour	3,200	31,500	12,000	157,500
Cooling Water, GPM	285	5,600	1,422	28,000
Labor, 24 Hr Day, Men	1	1	1	1.5
Capital Cost	$374,500	$2,386,500	$864,000	$6,270,000

- Means, and effectiveness of the means, of avoiding damage to the environment and the ecosphere as affected by the physical and chemical characteristics of the total waste stream

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*Pullman Kellogg Reference File number

The Solids Disposal Problem--

Production from coal of liquid or gaseous fuels containing 250 billion Btu per day will produce, as previously noted, on the order of 10.4 million to 25 million cubic feet per year of mixed solids consisting of ash with small amounts of inorganic salts, sludges, and catalysts. The volume may be reduced to some degree by compaction as the mass accumulates, but bulk densities cannot be expected above about 110 pounds per cubic foot for slag and about 60 pounds per cubic foot for non-slag ash.

Since the compacted slag and ash densities are in the general range of packed earth, there will be little difference in overall appearance between mounds made by depositing ash/slag directly on level ground and covering it with a layer of earth and mounds made by digging pits or trenches in level ground, filling them and then piling earth on top. For example, if the ash or slag for one year's production is deposited in a square area, 1000 feet on a side, and covered with 2 to 4 feet of earth, the mounds will range in height from 20 to 26 feet above grade. In 20 years the total area covered in a square configuration would be about 4500 feet on a side, amounting to about 465 acres. Obviously, piling higher would reduce the acreage, but against this must be considered the effect on the overall environment and the resident escophere. Final decisions on height and square versus rectangular or other plot areas can best be made on a site-specific basis that includes an environmental and an economic study.

Not the least of the problems involved in solid waste disposal is preventing leachates from contaminating the soil areas surrounding the disposal pile and the surface and ground waters. Interposing an impervious membrane between the wastes and the environment offers an attractive means of avoiding environmental damage. Such membranes will be discussed later.

Disposal of solid wastes in abandoned areas of strip mines appears to be an attractive alternative to piling the wastes at grade, particularly when it is considered that the volume of ash may range from 5 to 15 percent of the volume of coal fed to the conversion plant. While this method of disposal appears to be convenient, as when the mine is adjacent to the conversion plant, there are other considerations, such as:

- Means of impounding the solid wastes to prevent water erosion
- Means of access to the disposal area
- Means of preventing intrusion of ground water
- Means of preventing leachates from entering the environment

Construction of dams of earth or overburden is practical for impounding the solid wastes, provided that the other criteria can be met. Because the abandoned mine is a stripped-out area, the contours and the mine surfaces may be difficult to seal so that no intrusion of water or leakage of leachate will occur. Problems that maybe encountered are discussed in connection with FGD sludge disposal for Columbus and Southern Ohio Electric (1) and include the necessity for extensive regrading to obtain a sealed subsurface, the probability that leaching could not be prevented because of the presence of random piles of soil and rock overburden and the high cost of transportation of the stabilized sludge to the area. These same criteria must be applied to a feasibility study for solid waste disposal in strip mines.

Control of Leaching by Membranes--
Leachate from the mixed solids in the disposal site may be a portion of the free liquid phase of the solids as transported to the disposal site together with seepages of ground water through the site and drainage from precipitation. If the solids are in direct contact with the soil the possibility of percolation of the leachate into the soil under the site and movement both horizontally and vertically could eventually lead the leachate into subsurface waters. If the soil at the disposal site is relatively impervious, or the quantity of leachate is large, either continuously or intermittently, runoff may lead to contamination of surface waters.

It appears to be possible to isolate the solids from the soil by interposing an impervious membrane, or liner. With attention to grading, leachate can be conducted to a gathering sump from which it may be pumped back to the conversion plant for recycling as ash quench water or to treatment and preparation for use as process water. These conclusions are deduced from the results of investigations into use of liners in sanitary landfills, and the differences between sanitary landfill operation and conversion plant solid waste landfill operation:

- As previously noted, solid wastes contain no organic constituents. Sanitary wastes contain, among others, hydrocarbons, fats, animal and vegetable oils and, as bacterial action progresses, degradation products from these. The organic compounds have been shown in experiments to affect some types of disposal site liners (2)(4), resulting in failure, loss of strength or increases in permeability.

- Sanitary leachates have a pH around 5 (2)(4). Liner materials in both sanitary and inorganic waste service have been shown experimentally to be resistant to acidic waters at this pH (2)(3)(4). Trials of liners with nitric acid solutions showed (3) that cracking, hardening and surface blistering can be expected at low (1.5) pH, effects that may be partly caused by the oxidizing action of the nitric acid. A bentonite-sand admixture failed in the pH5 acid exposure while a soil-cement admixture apparently was resistant (3).

- Conversion plant ash usually has a pH around 8, varying with the coal composition. In short term tests, liner materials exposed to spent caustic at a pH of 11.3 showed no visible change, the bentonite-sand admixture failed and the soil-cement admixture appeared to be resistant to attack (3).

- Lined pits may be established in clay beds or by use of remolded clay (6). Consideration of the sorption and/or ion exchange characteristics of clays and the permeability of the prepared natural clay beds is essential. Permeability may be very low, on the order of 0.1×10^{-8} to 1×10^{-7} centimeters per second for distilled water, for clay landfill liners (6).

Asphalt Liners--In sandy soils a firm base for installation of the liner must be prepared by stabilizing the sand to a depth of 4 to 6 inches with asphalt emulsion. Specifications and instructions for application are contained in "Asphalt for Waste Water Retention in Fine-Sand Areas," published by The Asphalt Institute, and are included in Appendix D of Reference 5. Preparation of the lagoon area is of utmost importance, in that all debris, vegetation, and organic materials must be removed, areas to be paved must be graded and free of excess material and weak areas must be repaired. The sand layer and the asphalt are mixed by a travel plant, rotary or mechanical mixing, or by motor graders, then the mixture is spread evenly and compacted by conventional road machinery. Normal application of asphalt is 0.5 to 0.7 gallons per square yard per inch of compacted depth.

Where the soil is stable when dry, the asphalt stabilization may not be needed. The decisions are site-specific and are made by soil engineers. The precautions for site preparation are the same, in either case.

When the water depth (hydraulic pressure) is 8 feet or less an asphalt membrane seal is applied to the base over a tack coat. There are three allowable alternatives:

1. Hot-sprayed asphalt placed in 3 applications to a total of at least 1.5 gallons per square yard

2. Asphalt emulsion and 0.375 inch aggregate placed in 2 applications to a total of at least 0.5 to 0.7 gallons of asphalt per square yard plus 24 to 36 pounds of aggregate per square yard.

3. Hydraulic asphalt concrete placed in one course to a compacted thickness of 2 inches. The hot-mixed concrete, containing 6.5 to 9.5 percent by weight of asphalt with the balance 0.375 inch aggregate, is spread and compacted with conventional road building equipment.

When the water depth or hydraulic pressure is above 8 feet but not above the maximum allowable depth of 12 feet, hydraulic asphalt concrete is applied over a tack coat in two courses with overlapping joints to a total compacted depth of 3 inches.

The seals and linings must be extended beyond the crests of slopes as a means of anchoring the lining in the embankment roadway or top in order to prevent erosion damange.

It should be noted that use of hydraulic asphalt concrete as a liner allows operation of mechanical equipment in the area for spreading or compaction of the deposited solids when the lagoon impounds landfill solids.

Asphalt Sealed Fabric Liners--An example (5, Appendix E) of this type of membrane is nonwoven polypropylene fabric, fused one side, coated to a final thickness of 100 mils (0.1 inch) with two coats, totalling 1.4 gallons per square yard, of a mixture of:

Anionic asphalt emulsion SS-1h	100 gal
Asbestos fiber 7M-02	60 lb
Water	44 gal
Wetting agent (Phillips or equiv.)	2 lb

Joints in the fabric must be machine sewn. Edges are anchored in a perimeter trench and backfilled after sealing.

These specifications were formulated by the Soil Conservation Service (Engineering Standard 521-E-1), who also specified, among other characteristics, that the completed liner should have a Mullen hydraulic burst strength of 200 pounds per square inch and should exhibit no water loss when a hydraulic head of 35 feet of 100° F water is applied for 7 days.

Catalytically Blown Asphalt--This material is used as a seal for asphalt canal liners and to seal off layers of expansive soils under pavement (2). Prepared by air blowing hot asphalt in the presence of a catalyst (ferric chloride or phosphorus pentoxide), it is applied at 200 to 220° C in two applications totalling about 1 to 1.5 gallons per square yard and forms a film about 0.2 to 0.3 inches thick. The cooled membrane is flexible, tough, impervious to water and remains flexible at low temperatures. The membrane is usually covered with a layer of soil to protect it from traffic and from damage by light.

Flexible Membranes--Extreme care is required in site preparation so that the foundation area is smooth and free of projections that might damage the lining: stumps and roots must be removed and rocks and hard clods must be removed, rolled into a sub-base or covered with a fine soil cushion (Soil Conservation Service Standard S-521-A-1). Lining is spread smoothly and field spliced. The joints are required to develop a minimum of 60 to 80 percent of the film shear strength, depending on the film. A cover at least 6 inches thick of earth or earth and gravel (9 inches where the liner is exposed to livestock), with the bottom 3 inches no coarser than silty sand, is applied to protect the membrane.

Butyl rubber, chlorinated polyethylene, chlorosulfonated polyethylene, ethylene propylene rubber, polyethylene, and polyvinyl chloride are examples of materials that have been used to contain fluids, as in ponds. With the plastic films, a minimum thickness of 8 mils is required for application over sands and 12 mils for application over gravels. When rubber sheeting is reinforced with nylon the respective minimum thickness are 20 mils and 30 mils. Unreinforced rubber requires a minimum thickness of 30 mils for both soil types.

Soil-Cement Liners--Although mixtures of soil and cement have not been used as landfill liners, they have been in use for several years as paving materials and have shown excellent resistance to moisture penetration. Suggested specifications for the soil-cement base course have been assembled by the Portland Cement

Association (5, Appendix I). The soil must pass a 1-inch sieve and not more than 20 percent may be retained on a No. 4 sieve. The mixture may be prepared in place or in a central plant.

For in-place mixing, the cement, at 3 to 20 percent by weight of the soil, is spread uniformly on top of the soil and is mixed to a depth of about 6 inches with a disc harrow. Water is applied and mixed in and finally the whole mass is compacted and graded. Moisture retention for curing is accomplished by application of "bituminous material" applied at a rate of about 0.7 gallons per square yard. Curing is completed in about 7 days. According to the Portland Cement Association, the soil-cement course will withstand light local traffic immediately and all traffic in 7 days, "provided the soil-cement has hardened sufficiently to prevent marring or distortion of the surface."

Soil-Clay Membranes--The sodium variety of bentonite clay swells to up to 10 times its volume in water and, when mixed with a permeable soil, will reduce the permeability in direct proportion to the amount and type of bentonite added. The bentonite is distributed on the prepared soil base at 18 to 36 pounds per square yard and then mixed with the soil to a depth of 4 to 6 inches with a disc harrow. The soil mixture, after compaction, is ready for use.

Because bentonite, like other clays, exhibits ion exchange properties, the composition of the leachate in contact with it may affect its sealing properties. For example, calcium bentonite does not swell as does sodium bentonite. Sodium bentonite will readily exchange its ions for the calcium ions in solution and become less effective as a sealant (7).

A proprietary mixture of sodium bentonite and polymers that swell in water is more resistant to the effects of ion exchange than bentonite alone (8). The material is applied in the same manner as bentonite alone, but normally at a rate of 9 to 18 pounds per square yard. Applications noted by the manufacturer include sanitary landfills, liners under stockpiles and mine refuse piles and sewage and process water lagoons.

Evaluation of Liners--

In an ongoing EPA study (4) a variety of liner materials were exposed to leachate from simulated sanitary landfills. The study was undertaken with these objectives:

- To determine the effects of exposure to leachate from compacted municipal refuse on the physical properties of lining materials (excluding soils and clays) that are believed to be potentially useful for the lining of sanitary landfills

- To estimate the effective life of liner materials when exposed to prolonged contact with leachate under conditions comparable to those encountered in a sanitary landfill

- To determine the effects of exposure for 12 to 24 months to sanitary landfill leachate on the physical properties of the 12 liner materials mounted in the bases of the simulated sanitary landfills and on the 42 smaller specimens buried in the sand placed above the mounted liners

- To analyze the costs of these materials for lining sanitary landfills. This analysis will include liner costs, installation costs, and the benefits from longer durability

The lining materials tested were:

Hydraulic asphalt concrete (3 percent voids)
Paving asphalt concrete (6 percent voids)
Soil asphalt
Soil cement
Blown asphalt (canal lining asphalt)
Emulsified asphalt on fabric
Butyl rubber
Chlorinated polyethylene (CPE)
Chlorosulfonated polyethylene (Hypalon)
Ethylene propylene rubber (EPDM)
Polyethylene (PE)
Polyvinyl chloride (PVC)

After a year of exposure to the leachate from the simulated landfill there were these observations:

- The admix liners containing asphalt, although losing drastically in their compressive strength, maintain their impermeability to leachate. The asphalt itself became softer, indicating possible adsorption of organic components from the leachate

- During the year's monitoring of the cells, in only three of the cells did the leachate enter the base below the liners. Two of these liners, soil asphalt and paving asphalt concrete, leaked. The leakage in the third was caused by a failure of the epoxy sealing compound around the periphery of the specimen

- The soil cement lost some of its compressive strength; however, it hardened considerably during the exposure period and cored like a Portland cement concrete. Its permeability decreased somewhat

- Inhomogeneities in the admix materials, which probably caused the leakage in the paving asphalt and soil asphalt liners, indicate the need for considerably thicker materials in practice (2 to 4 inch thick liners were selected for this experiment to give an accelerated test and were designed with an appropriately sized aggregate)

- The asphaltic membranes withstood the leachate for 1 year, although they did swell slightly. There was no indication of disintegration or dissolving of the asphalt

- All of the polymeric liner materials withstood a 1-year exposure to the leachate, although several, e.g. chlorinated polyethylene and Hypalon, swelled appreciably. Swollen liners softened but did not lose tensile, tear, or puncture resistance. Preliminary tests of the exposed liners indicated some increase in permeability, probably because of swelling.

- Variation occurred among polymeric membrane liners based upon a given polymer, which may reflect variations in polymer source, compound composition, and possibly methods of manufacture.

- The seams of the polyvinyl chloride, Hypalon and chlorinated polyethylene liners deteriorated in strength. The polyethylene retained its strength best.

- The quality of the leachate in all 24 of the cells was similar, indicating that the initial composition of the refuse was controlled and that the comparison among the liner materials would be valid.

The average pH of the leachate was 5.1. The COD averaged 46 grams per liter and the volatile acids (acetic, propionic, butyric, isobutyric) averaged 22 grams per liter.

The overall conclusion may be drawn that the deterioration in physical properties that was noted in the asphaltic liners and in some of the polymers may have been due, at least in part, to the presence of organic materials in the leachate. It would be logical to suppose that these same liners, if exposed to the conditions in the solid waste disposal area of a coal conversion plant, where there are no organic materials, would exhibit less of a deterioration of physical properties. As a consequence, selection of lining materials for solid waste areas may be made primarily on the basis of such parameters as initial cost, cost of installation, ease of installation (seaming methods, special construction at joints), resistance to puncture or tearing, and resistance to oxidation and sunlight. In all cases the base, or subgrade, must be carefully prepared.

Choosing a Liner--

Following are some of the criteria that must be considered in choosing liner materials:

- Clay soils, including bentonites, can lose their impermeability when impounding strong acids, strong bases, or brines. Testing and careful evaluation may be required for selection of a clay that will remain impermeable

- Exposure of liner to sunlight. Some polymers, such as polyvinyl chloride, may become brittle and crack, either by loss of plasticizer by evaporation or by degradation by ultraviolet light. Special plasticizers and/or inhibitors may be required in polymer formulations. Butyl rubber is susceptible to cracking from ozone attack.

- Exposure of liner to weather. Soil liners may be damaged by drying or by freeze-thaw of exposed areas.

- Variability in the strength and chemical resistance of field splices of the polymeric materials. Polyethylene may be heat sealed or (usually) solvent sealed, as are Hypalon, chlorinated polyethylene and polyvinyl chloride. Tetrahydrofuran showed promise in the splicing of polyvinyl chloride(4). There are variations in the cements or solvent systems for the various polymers. In general, splices with solvent washes are difficult to make in the field (4)

Liner Protection--

None of the proposed liner materials should be used directly as a pavement. While some of the materials can easily support rubber-tired construction equipment, no manufacturer recommends allowing distribution vehicles to use the liner as a pavement because of the high wheel loadings. Equipment with crawler treads should not be allowed to operate directly on the liner. Manufacturers recommend protecting the liner with an earth cover one to two feet thick. This material should not contain jagged rocks or other sharp objects that could damage the liner. Similarly, the first lift of solid waste placed in the fill site should be placed carefully and spread so that the depth of the cover protecting the liner is effectively increased.

Control of Leaching by Chemical Stabilization--

Processes for chemical stabilization of waste materials have been receiving increasing attention due to the requirement for substitution of coal for natural gas and oil in power generation and the requirement for low sulfur dioxide emissions to the atmosphere. The public utilities appear to adhere to the principle that the principal effort should be generation of power and that there should be minimum effort required in handling emissions. The result of this attitude seems to be lack of interest in recovering sulfur from stack gases for sale and considerable interest in scrubbing the sulfur dioxide from the stack gases and disposing of the resulting sludge.

The volume of the flue gas desulfurization (FGD) sludge is expected to be large and the physical and chemical characteristics are poor (9). Processes have been developed for chemical stabilization of the sludge so that the soluble matter is immobilized, the permeability is reduced to a level comparable to that of the more impervious soils and the physical bearing strength is increased from an extremely low level to one resembling well-compacted earth. Although disposal of FGD sludge is not a real part of the present study, the results of the work done by various organizations in stabilizing the sludge may be applied toward solution of the general problem of solid waste disposal. Further, as pointed out earlier, the decision to recover sulfur or dispose of it will be site specific and will be a management choice based on the economics of the situation. Consequently, the solid waste from coal conversion plants may include varying amounts of FGD sludge, ranging from zero upwards. Since these sludges in general are thixotropic and rarely contain more than 50 percent solids, mixing them with the other solid wastes may, depending on the proportion of sludge, reduce the bearing strength of the total mixture to such a low level that the solids waste area cannot be reclaimed and may be a hazard.

Chemical stabilization has been applied to FGD sludge to improve its physical characteristics. The final result of the stabilization process may resemble a cement-like monolith of very high density. Permeability, of the solid mass is usually less than 10^{-6} cm/sec. Chemical stabilization appears to offer excellent possibilities for economical disposal of ash/slag alone, ash/slag plus FGD sludge or FGD sludge alone. Where both ash/slag and FGD sludge must be disposed of, an economic study may demonstrate advantages in two separate handling and disposal systems, one to dispose of the ash without stabilization and one to stabilize and dispose of the FGD sludge.

There are three chemical stabilization systems in commercial use: the Dravo Corporation Synearth process, the IU Conversion Systems (IUCS) Poz-O-Tec process and the Chemfix (division of the Carbonundum Company) treatment process. In the Dravo and IUCS processes there are cementitious reactions involving calcium, aluminum, and silicon compounds in the wastes and in the additives. In the Chemfix process, sodium silicate and Portland cement are added to the wastes and react with each other and with the wastes.

The Dravo Process--In the sequence of operations, the sludge is first dewatered to about 35 percent solids. "Calcilox" (a furnace slag) is added at about 7 percent of the dry sludge solids together with lime at about 2 percent of the dry sludge solids. Actual amounts may vary, depending on sludge characteristics. The slurry is pumped to a disposal pond where the solids settle, reaction ensues and the solids harden, in about 30 to 45 days, to a mass having an unconfined compression strength of around 4,000 pounds per square foot in 30 days to 6,000 or more pounds per square foot in 45 days (10). According to Dravo reports (11), the sludge mixture will dewater to about 45 to 50 percent solids.

The Dravo process is also being used to stabilize fine coal refuse (12). The final stabilized mixture at 42 percent solids showed compression strengths of about 1,200 to 1,700 pounds per square foot after 30 days, considerably less than the stabilized FGD sludge. In appearance, the stabilized material resembled a fine silt. The undisturbed material had a permeability ranging from 2 to 9 x 10^{-7} centimeters per second, depending on the proportions of Calcilox and lime mixed into the refuse.

The concept of operation of the process for solids disposal involves preparation of a pumpable slurry of the solid waste and the reactants and pumping to a dammed or diked area for settling, solidification and partial dewatering. The supernatant liquid is returned to the FGD scrubber or to the coal cleaning plant for reuse. When the disposal area is full of solids it may either be pumped dry and covered with earth for area reclamation or left as a lake. Because of the high pH of the stabilizing process it would appear that the advisability and the acceptability of the latter alternative is highly questionable.

The IUCS Process--Stabilization of fly ash, FGD sludge and mine refuse, alone and in combinations, have been successfully demonstrated on a large scale (13). Laboratory and pilot plant work has been correlated with the full scale results. Controlled mixing of the wastes and such additives as lime and, for FGD sludge, flyash is prescribed from laboratory investigations on the specific waste material.

With FGD sludge, lime, and flyash are added and mixed in a pugmill at about 60 percent solids (1). The damp mass is discharged to a surge pile and allowed to set for 3 to 6 days and is then transported, at a consistency similar to clay, to the landfill area, where it is distributed and compacted. The mixture develops a compressive strength that can range from around 10,000 pounds per square foot (1) upward to 20,000 pounds per square foot and higher, depending on the end use of the disposal area (13). Permeability is on the order of 1 x 10^{-6} centimeters per second or less (13). The stabilized material has been used to line a pond by applying successive layers each about 6 inches thick, and has shown a permeability in this service of about 1 x 10^{-7} centimeters per second.

Coal waste has been successfully stabilized by the IUCS system micrometers. In the several mixes evaluated in the laboratory, solids content of the stabilized material ranged from 77 to 81 percent, compression strength after 28 days ranged from 3,000 to 15,000 pounds per square foot and permeability ranged from 1 to 3 x 10^{-6} centimeters per second. The ranges in properties resulted from varying the proportions of additives to the mix.

Chemfix--For FGD sludge stabilization the dewatered sludge is mixed with sodium silicate and Portland cement. The silicate forms insoluble compounds with polyvalent metal ions, then reacts with the cement, while the cement hydrates and also reacts with sludge components (15). Proportions of the additive are varied, depending on the final characteristics desired in the processed material. For FGD sludge the sodium silicate usage is about 4 percent and the cement usage is about 7 percent of the weight of dry sludge solids (16).

No data are available from Chemfix on test results on stabilized FGD sludge or on mine refuse. Compressive strengths of various stabilized wastes from other sources ranged from 1,000 to 10,000 pounds per square foot. No permeability data were available.

In an economic study of FGD solids disposal, conducted by TVA (16), the Chemfix process was compared to the IUCS and Dravo processes. The system used in the study mixed the sodium silicate and the cement with FGD slurry that had been dewatered to 60 percent solids. The mixture is then trucked to the disposal site and compacted.

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COST OF SOLIDS DISPOSAL

Preparation of a Lined Disposal Site

The annual volume of solids produced in coal conversion processes was estimated previously to be 10.4 to 25 million cubic feet. To contain this volume of solids, and to provide means of collection of any runoff waters, the mass should be confined. One method of preparing a relatively flat area for a disposal site is shown in Figure 10-2.

In Figure 10-2 the configuration of the disposal area was developed from the following:

1. The earth removed below grade shall be used to construct the walls
2. The wall slope shall be 3 horizontal to 1 vertical
3. The top of the walls shall be a roadway 20 feet wide
4. The disposal area shall be square in order to provide minimum perimeter with maximum interior surface area
5. The total volume shall be 110 percent of the solid wastes volume

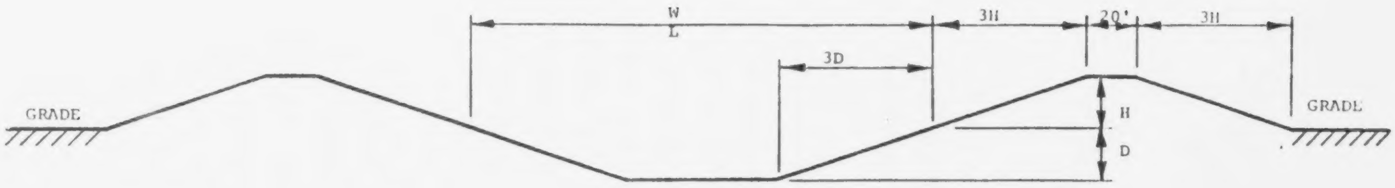
GRADE
W
L
3D
3H
20'
3H
H
D
GRADE

Figure 10-2. Solid wastes disposal area.

The volume below grade = $W^2D - 4(3WD^2 - 9D^3)$

where W = Width at grade = Length at grade

D = Depth below grade

The total impounded volume is:

$$W^2H + 6\ WH^2 + 12\ H^3 + W^2D - 12\ WD^2 + 36\ D^3$$

where H = Height above grade

The volume of the wall above grade =

$$4(3\ WH^2 + 20\ WH + 18\ H^3 + 180\ H^2 + 400\ H)$$

where H = Height above grade

From these relationships the volume of earth to be moved and the area of liner required may be calculated for various lengths (widths). Results of the calculations are shown in TABLE 10-8.

In TABLE 10-9 are shown installed costs of various lining materials for disposal areas. The two opinions of Haxo and Clark on the installed costs have been updated to the 1977 basis for comparison. There is considerable difference between the two, with Haxo generally being higher. Private communications with several manufacturers and installers of the pond liners indicate that the Clark film prices are probably representative, while for the asphalts, bentonite,and the soil-cement liners the Haxo prices are probably representative.

There are several possibilities for collecting the drainage and runoff from solid waste impoundment areas. One method would involve placing a layer of washed sand on the liner and topping the sand with gravel. Drainage pipes through the walls of the impoundment area would be manifolded or would discharge into a peripheral trench which would conduct the drainage to a central sump. Waters collected in the sump would be pumped back to the

TABLE 10-8. ESTIMATED DISPOSAL SITE CONSTRUCTION COSTS

A. Site for Disposal of 25 Million Cubic Feet of Solids

L Length	W Width	H Height Above Grade	D Depth Below Grade	Earth to be Moved	Earth Moving Cost (2)	Internal Surface Area (3)	Plan Area, Acres
1250 ft.	1250 ft.	14 ft	3 ft	168,600 CY(1)	$202,300	199,300 SY (1)	43
1000	1000	19	7	237,900	285,500	139,800	30
750	750	26	29	356,40	427,700	94,200	21

B. Site for Disposal of 10.4 Million Cubic Feet of Solids

L Length	W Width	H Height Above Grade	D Depth Below Grade	Earth to be Moved	Earth Moving Cost	Internal Surface Area (3)	Plan Area, Acres
1000 ft	1000 ft	9 ft	2 ft	33,700 CY	$ 40,000	124,300 SY	27
750	750	14	5	62,100	74,500	78,400	17
600	600	18	14	108,600	130,300	57,100	9

(1) CY = Cubic yards, SY = Square yards

(2) Cost = $1.05 per cubic yard for earth moving plus $0.15 per cubic yard for compaction and final grading (mid 1978)

(3) Area to be covered or treated for reducing or eliminating permeability

TABLE 10-9. INSTALLED COSTS OF DISPOSAL AREA LINERS

	Installed Cost per square yard(1)	
	Haxo (2)	Clark (3)
Polyethylene, 10-30 mil	$1.27 - 2.04	
Polyvinyl chloride, 10 mil		$1.07
20 mil		1.63
30 mil		1.95
10-30 mil	1.66 - 3.06	
Chlorinated polyethylene, 20 mil		2.89
30 mil		3.78
20-30 mil	3.44 - 4.59	
Hypalon, 20 mil		2.89
30 mil		3.78
20-45 mil	4.08 - 4.33	
Butyl rubber, 31.3 mil		3.33
46.9 mil		4.00
62.5 mil		4.60
31.3-62.5 mil	4.60 - 5.66	
Ethylene-propylene diene monomer, 31.3 mil		3.22
46.9 mil		3.89
62.5 mil		4.55
31.3-62.5 mil	3.44 - 4.84	
Paving asphalt + sealer, 2 in.	1.70 - 2.41	2.22
4 in.	3.33 - 4.60	
Hot Sprayed asphalt, 1 gal/sq. yd.	1.74 - 2.45	
Asphalt membrane, 100 mil	1.78 - 2.65	1.55
Soil-bentonite, 9 lb/sq. yd.	1.02	
18 lb/sq. yd.	1.66	1.55
Soil-cement + sealer, 6 in.	1.77	

(1) Includes material and labor. Does not include cost of subgrade preparation or earth cover

(2) From Reference 5, p.17, updated from 1973 to 1977 by Chemical Engineering Cost Index

(3) From Reference 5, p.18, updated from 1974 to 1977 by Chemical Engineering Cost Index

conversion plant for reuse as ash quench water or, following treatment, as process water. Another method of collecting drainage would be to grade the bottom of the impoundment to a central sump from which a pipe or channel would carry the drainage to a pumping station. Of the two methods, the former would probably be preferred because all piping would be accessible for inspection and maintenance. Costs for the drainage system, not including sump, pumps or pipelines to the conversion plant, are estimated to be on the order of \$0.75 to \$0.90 per square yard of impoundment surface area.

Comparison of the estimated capital costs of impoundment areas of various dimensions may be made with the following criteria:

- Land cost at \$3,500 per acre
- Earth moving cost as in TABLE 10-8
- Drainage system at a median cost of \$0.83 per square yard
- Liner costs from TABLE 10-9

These costs are summarized in TABLE 10-10 for the two quantities of solids that are under consideration. It is seen that in both Case A and Case B the advantage lies with the smaller, deeper impoundment areas. The advantage is particularly apparent when the total disposal area for the life of the conversion plant, which may be 20 to 30 years, is considered.

Disposal Sites with Soil and Clay Lining

It is generally stated by soil engineers that soils with a permeability rate of 10^{-6} centimeters per second or less may be considered as being impermeable since, at this rate, migration of liquid through such a soil would be on the order of 12.5 inches per year. The most stringent of environmental regulations

TABLE 10-10. ESTIMATED CAPITAL COSTS OF LINED DISPOSAL AREAS

A. Disposal of 25 Million Cubic Feet of Solids

	Impoundment Dimensions, Feet		
	1250x1250	1000x1000	750x750
Land Cost	$150,500	$105,000	$ 73,500
Earth moving cost	202,300	285,500	417,700
Drainage system cost	144,100	92,200	51,900
Liner cost: high (1)	797,200	559,200	376,800
medium (2)	566,000	397,000	267,500
low (3)	352,800	247,500	166,700
Finish cover installation (4)	138,900	88,900	50,000
Total, high cost liner (1)	$1,433,000	$1,130,800	$979,900
medium cost liner (2)	$1,201,800	$ 968,600	$870,600
low cost liner (3)	$ 988,600	$ 819,100	$769,800

B. Disposal of 10.4 Million Cubic Feet of Solids

	Impoundment Dimension, Feet		
	1000x1000	750x705	600x600
Land cost	$ 94,500	$ 59,500	$ 31,500
Earth moving cost	40,400	74,500	130,300
Drainage system cost	92,200	51,900	23,100
Liner cost: high (1)	497,200	313,600	228,400
medium (2)	353,000	222,700	162,200
low (3)	220,000	138,800	101,100
Final cover installation (4)	88,900	50,000	38,000
Total, high cost liner (1)	$813,200	$549,500	$451,300
medium cost liner (2)	$669,000	$458,600	$385,100
low cost liner (3)	$536,000	$374,700	$324,000

(1) Paving asphalt + sealer, 4 inch thickness, $3.33 to $4.60 per square yard, median for calculation = $4.00
(2) Average of costs in TABLE 10-9 at $2.84 per square yard
(3) Soil cement + sealer, 6 inches thick, $1.77 per square yard
(4) Earth cover, 2 feet thick, installed and compacted at $0.80 per square yard

(Dominion of Canada) requires this same maximum permeability rate. It is possible to achieve this permeability by compaction of suitable soils, admixture of native clay and remolding of native clay. Aside from the difficulty of creating a homogeneous barrier layer, clays have been shown to be susceptible to ion exchange, with consequent changes in their physical characteristics, in contact with leachate from sanitary landfills. Much work has been done in evaluating barriers for handling FGD sludges. Ponds have been in commercial operation for many years for storing flyash slurry from power plants and fine coal and refuse from mining operations. None of the storage or impoundment conditions, however, duplicate those to be found in impoundment of gasifier ash. Judgement as to the suitability of native soils and clays, whether virgin, remolded or compacted must, therefore, be by inference.

Cost of preparing a clay lined site may be estimated from the costs of preparing the base for membrane liner and then adding the following as appropriate:

- Clay for the lining assumed delivered to the site at $1.00 per cubic yard
- Spreading and compacting assumed at $1.20 per cubic yard
- Liner to be 3 feet thick
- If native soil is suitable, assume $0.60 per cubic yard for spreading, compacting, and final grading.

Estimated costs are shown in TABLE 10-11.

From the foregoing discussion it is obvious that despite the apparent economics to be realized through use of available clays

TABLE 10-11. ESTIMATED CAPITAL COSTS OF SOIL AND CLAY LINED DISPOSAL AREAS

A. Disposal of 25 Million Cubic Feet of Solids

	Impoundment Dimensions, Feet		
	1250x1250	1000x1000	750x750
Cost of Land, earthmoving, drainage system, and final cover	$551,300	$463,700	$444,800
Liner cost: clay	442,400	314,200	215,400
native soil	120,700	85,700	58,700
Total: clay liner	$993,700	$777,900	$660,200
native soil liner	$672,000	$459,400	$503,500

B. Disposal of 10.4 Million Cubic Feet of Solids

	Impoundment Dimensions, Feet		
	1000x1000	750x750	500x500
Cost of land, earthmoving, drainage system and final cover	$316,000	$235,900	$210,600
Liner cost: clay	273,200	174,700	104,500
native soil	74,500	47,600	28,500
Total: clay liner	$589,200	$410,600	$315,100
native soil liner	$390,500	$283,500	$239,100

and compacted native soils, experimentation to determine suitability for the service is a necessity.

Disposal Sites with Chemically Stabilized Solids

Published information on chemical stabilization does not include stabilization of ash alone. Much has been published on stabilization of FGD sludge and mine refuse, and the following discussion and cost development is based on this information by inference.

To determine the characteristics of stabilized ash/slag solids with regard to permeability and strength in compression and shear, to reach decisions on the suitability of chemical stabilization for this service, and to determine the economics of the systems, experimentation on actual materials is needed.

Leo, Fling and Rosoff (1) develop summarized costs for stabilization and landfilling of FGD sludge containing 40 percent flyash from a 500 megawatt power generation plant over a 30 year life. The land used was 320 acres, or about 11 acres per year, with sludge piled 30 feet high. Land price was \$5000/per acre. Dry solids in the sludge generated totalled 250,000 tons per year. A previous estimate of total disposal costs considered the Dravo, IUCS and Chemfix stabilization systems. No distinction between processes is made in the statement that, in 1977, the total disposal costs were 1.53 mills per kilowatt hour or \$12.27 per ton of dry solids. (It should be noted that the sludge disposal cost calculated from the plant capacity and the stated cost per kilowatt hour is:

$$500{,}000 \times 4{,}380 \times \$0.00153/250{,}000 = \$13.40 \text{ per ton of dry sludge solids.)}$$

This cost includes capital and operating costs for all facilities, chemicals, utilities, and manpower required for the preparation and disposal of the FGD sludge.

In a TVA study (2) the Dravo, IUCS and Chemfix processes were compared in systems for disposal of about 250,000 tons per year of FGD sludge containing flyash from a 500 megawatt power station operating at conditions similar to the previous case. In this study a midwestern plant location was selected. Land costs were assumed to be $3,500 per acre. Economic assumptions were as follows:

- All capital cost estimates based on Chemical Engineering cost indexes (labor index--237.9, material index--264.9). Capital costs projected to mid-1979. Project assumed to start in mid-1977 and be completed in mid-1980.

- Direct capital costs covered process equipment, piping and insulation, transport lines, foundations and structural, excavation and site preparation, roads and railroads, electrical, instrumentation, buildings, pond construction, and earthmoving equipment. Material and labor (fabrication and installation) costs for each of these items were estimated.

- Indirect capital costs included engineering design and supervision, architect and engineering contractor expenses, construction expenses, contractor fees, contingency, allowance for startup and modifications, and interest during construction. Working capital and land were included in the total capital requirements. Estimates were based on current industry practice and authoritative literature sources.

For the Dravo system, the thickened sludge at 35 percent solids was treated with Dravo additives: Calcilox at 7 percent of dry sludge, and Thiosorbic lime at 2 percent of dry solids. A gravity thickener was used for dewatering the sludge. Treated sludge was pumped to a clay-lined pond located about 1 mile from the scrubber facilities. The ponded sludge was assumed to settle in the pond to 50 percent solids, with excess water recycling to the scrubber system. The treated settled sludge fixed as a soillike material in the pond.

For the IUCS system, the effluent from the scrubber system was dewatered to 60 percent solids using a thickener and rotary drum filter. The dewatered sludge was fixed by mixing with lime at 4 percent of dry sludge solids. Trucks transported the treated sludge to the landfill disposal site located 1 mile from the scrubber facilities. The treated material was assumed to have claylike properties that allowed placement and compaction in the landfill with earthmoving equipment.

For the Chemfix system the effluent from the scrubber system was dewatered to 60 percent solids using a thickener and rotary drum filter. Dewatered sludge was fixed by mixing with Chemfix additives: Portland cement, at 7 percent of dry solids, and sodium silicate, at 4 percent of dry solids. The thickened sludge, at 35 percent solids, was pumped 1 mile to the disposal site where it was filtered to remove additional water and mixed with additives. The treated material was then hauled to the landfill, placed, and compacted with typical earthmoving equipment.

The capital investment requirements for the three processes were estimated to be:

	Capital Investment, $1000		
	IUCS	Chemfix	Dravo
Direct costs(1)	$ 4,301	$ 5,775	$ 4,943
Other direct costs(2)	581	442	7,410
Indirect costs(3)	2,955	3,700	6,381
Startup and interest(4)	1,666	2,138	3,381
Land (5)	676	693	1,450
Working capital	538	783	550
Total capital investment	$ 10,717	$ 13,531	$ 24,115

(1) Does not include trucks, earthmoving equipment, or pond

(2) Trucks and earthmoving equipment for IUCS and Chemfix, pond for Dravo

(3) Includes 20 percent contingency on engineering, procurement, construction and contractor's fee

(4) Allowance for startup and modification and for interest during construction

(5) 193 acres for IUCS, 198 acres for Chemfix, 414 acres for Dravo

The TVA study concludes that the unit revenue requirements for sludge disposal are:

	Per Ton of Dry Solids	Mills/KWH
IUCS	$12.55	1.51
Chemfix	$16.51	2.00
Dravo	$15.32	1.91

Thus, IUCS has an apparent advantage over the others in both capital and operating costs, while Chemfix has a lower capital cost than Dravo but a higher operating cost.

In a report (3) on the operation of the Bruce Mansfield power plant, the capital cost (1975) is presented as $90 million for a complete Dravo FGD disposal system for 3 power units generating a total of2,475 megawatts. Escalating to 1977 and factoring downward to 500 megawatt capacity yields an estimated capital investment of $36 million, which may be compared to the TVA estimate of $24 million with these modifiers:

- At Bruce Mansfield the slurry is transported 7 miles, as against TVA's assumption of 1 mile
- At Bruce Mansfield the impoundment area is 1,400 acres and includes a dam that will eventually reach a height of 420 feet, 2,200 feet long at the crest and 1,550 feet thick at the base. Although the proportion of the total capital represented by the impoundment area is not stated, it would appear to be substantial and account for much of the factored difference in capital cost

IUCS has applied their Poz-O-Tec process to stabilization of a mixture of coal mine refuse and FGD sludge (4). In the laboratory work the mixture stabilized satisfactorily and exhibited permeabilities of 1×10^{-6} to 3×10^{-6} centimeters per second. No costs for this operation were developed. IUCS states that for FGD sludge the typical cost may range from $3 to $7 per ton of dry solids. The basis for these costs was not stated, and they are apparently not on the same basis as those developed by the TVA study.

In another report of operation of the IUCS system for disposal of FGD sludge (5) from 1,625 megawatts of generating capacity costs are developed for sludge disposal onsite in a system engineered, built and operated by IUCS. About 304,000 tons per year of solids (dry basis) are impounded in an area of 50 acres. Because of the demonstrated load bearing characteristics of the stablilized sludge, at over 10,000 pounds per square foot, construction of a 100 foot mound over a period of 20 years appears to be feasible. The electric company pays an annual fee to IUCS and supplies operating and maintenance labor, heavy equipment maintenance, auxiliary power and lime when the demand is over 3.5 percent of the dry solids processed. On this basis, the total operating cost to the elctric company is $10.78 per ton of dry solids when the plant load factor is 51 percent. At a 70 percent load factor the conversion system processes about 416,000 tons a year and, because the IUCS fee and operating labor are constant, the total operating cost drops to $8.53 per ton of dry solids. When these cost figures are prorated to the processing of 250,000 tons per year of dry solids, the total operating cost becomes $12.60 per ton of dry solids, showing very close agreement with the TVA study.

The conclusions that may be drawn from this information are as follows:

- Chemical stabilization appears to offer great promise as a means of economically disposing of solid wastes without endangering the environment

- The IUCS stabilization process has the lowest capital and operating costs of the three processes studied by TVA

- The results of the TVA study of process operating costs for the IUCS process are borne out by commercial operation

- The available useful information on stabilization of solid wastes concerns FGD sludge disposal

- Operation of the stabilization processes with the solid wastes from coal conversion plants may differ from operation with FGD sludge and may require different process operating parameters and changes in the types or quantities of additives, which in turn may affect capital and operating costs

- Laboratory and large scale testing of stabilization processes is needed to determine the permeability and compressive strengths of stabilized solid wastes from conversion processes.

- Results of the permeability tests will aid in determining whether or not the disposal area must have a lining to prevent leachates from reaching ground and surface waters

- Economic studies of the three stabilization processes that have been discussed, and others that show promise, are needed to establish the similarities and differences between processes and between operation with FGD sludge and conversion plant solids

If the assumption is made that data from FGD sludge stabilization with the IUCS system may be applied to coal conversion solid wastes, then estimates may be made of stabilization costs for the solids from gasification of low and high sulfur coal and from liquefaction of high sulfur coal. These annual operating costs and the cost per dry ton of solids processed are shown in TABLE 10-12. The advantage of size of operation, in comparison to the previously-developed costs for FGD sludge disposal, is immediately apparent.

References

1. Leo, P. P., Fling, R. B., and Rosoff, J., "Flue Gas Desulfurization Waste Disposal Field Study at the Shawnee Power Station." Presented at Flue Gas Desulfurization Symposium. Florida, November 1977. 746*

*Pullman Kellogg Reference File number

TABLE 10-12. ESTIMATED ANNUAL OPERATING COSTS FOR STABILIZATION OF COAL CONVERSION PLANT SOLID WASTES WITH THE IUCS PROCESS

	Gasification		Liquefaction
	Low Sulfur	High Sulfur	High Sulfur
Dry solids, TPY	518,400	616,800	700,900
IUCS annual fee	$2,384,000	$2,384,000	$2,384,000
Operating labor	140,000	140,000	140,000
Maintenance labor	87,000	96,000	104,000
Heavy equipment maintenance	174,000	207,000	236,000
Auxiliary power	159,000	176,000	190,000
Lime usage	800,000	949,000	1,079,000
Total annual cost	$3,744,000	$3,952,000	$4,133,000
Cost per ton of dry solids	$7.22	$6.41	$5.90

Note: Costs prorated from data in Reference 5

2. Barrier, J. W., Faucett, H. L., and Henson, L. J., "Economics of FGD Disposal." Presented at Flue Gas Desulfurization Symposium. Florida, November 1977. 751*

3. Lobdell, L. W., Rothfuss, E. H., Jr., and Workman, K. H., "Eighteen Months of Operation (of the) Waste Disposal System (at the) Bruce Mansfield Power Plant, Pennsylvania Power Company." Presented at Flue Gas Desulfurization Symposium. Florida, November 1977. 748*

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NEED FOR FURTHER STUDY

Coal Dust Control

Conversion of power generation plants to coal firing, construction of new coal fired generation facilities and the establishment of coal conversion processes all increase the movement of coal from mines to points of use. Since it is unlikely that coal users will always be located within conveyor distance of coal producers, rail transport of coal may increase by an order of magnitude by 1985 to 1990. Dust problems arising from coal transportation may be expected to increase in proportion. There appears to be a need for studies directed toward dust control or dust elimination, with emphasis on the latter.

Coal fines have been treated as a necessary evil in coal cleaning. Creation of fines begins at the seam face and continues through the crushing and screening operations. Reduction in the fines content of the coal broken from the seam face may be possible by modification of the tooth profiles of continuous miners, the angle of attack of the teeth against the face or the tooth velocity at impact. There is probably little that can be done to minimize fines creation during strip mining operations, since the coal must be broken from horizontal seams, instead of vertical faces as in underground mining.

As pointed out in delineation of the coal dust problem, crusher design and operation have a significant effect on fines creation. An investigation into the effect on dust creation of such operating parameters as peripheral speed, angle of attack of the crusher mechanism against the coal, tooth or jaw design and feeding method might lead to methods of operation or machine design that would reduce fines, decrease power requirements, and decrease machine costs.

Separation of the fines and dust from the main coal stream may be accomplished by dry or wet screening or by elutriation of dry fines by an air stream. In most coal preparation plant flowsheets the initial breaking of lumps to separate coal from refuse is a dry operation and dust created at this point may be collected in baghouses for later combination with wet fines streams. The total dust stream, however, may contain significant quantities of undesirable materials, such as clay and silt. In this case, separation of the non-combustible substances may be accomplished by air elutriation by taking advantage of differences in density of coal and non-coal particles.

Following the breaker, the coal stream is usually wet screened and separated into various size fractions from which refuse is to be separated. Attention to the means of applying water to the coarse and intermediate size coal streams during their residence on the screens could ensure that all fines are washed from the coal and sent to the fine coal cleaning section. Normal procedure for treating coal fines after separation from refuse involves combining the fines stream with the intermediate size stream and drying the mixture. The coal fines are thus recombined with the main coal stream.

A study of the mechanics and economics of compacting or agglomerating the fine coal stream to larger, non-dusting sizes, could yield attractive processes for eliminating dust and, possibly, for increasing the value of the coal: it must be rememberd that eliminating the dust at the source eliminates the need for dust control methods during transport, storage and handling at the mine and at the conversion plant site, and that these savings should defray part or all of the cost of dust elimination.

Solids Waste Disposal

As has been pointed out, the investigations on the applicability of impermeable liners for disposal areas has been confined to sanitary landfill application and, to a much lesser extent, to FGD sludge applications. Because the characteristics of ash and slag, and mixtures of these with other solids, from coal conversion processes are quite different from those of sanitary landfills, FGD sludges or power plant ash, conclusions reached in the fines testing programs that are now in operation cannot be applied with confidence to coal conversion solid wastes. A program for evaluation of liners in coal conversion waste disposal service would be a most valuable aid in establishing applicability of liners and economics of their use. This information may become of prime inportance in the future if environmental standards that concern permeation of the soil by leachates become more stringent.

With regard to chemical stabilization of solids, for the three systems (IUCS, Dravo, and Chemfix) that have been discussed and others that are as yet untried on a large scale, the principal application has been in stabilizing FGD sludge. Successful as this application has been, the parameters of chemical stabilization of FGD sludge cannot be applied immediately to solid wastes from coal conversion processes. For example, the ash/slag may be either easier or more difficult to stabilize, entailing more or less additive usage, special handling procedures or special waste are a management procedures, any of which may affect the economic attractiveness of the methods. Further, the permeability of stabilized ash/slag may differ widely from that of, say, FGD sludge: it may be greater, and thus possibly unacceptable by present standards, or it may be significantly less and thus offer an excellent possibility for solid waste disposal that will meet present standards and possible future environmental goals. A testing program to determine the physical, chemical, and economic aspects of chemical stabilization appears to be necessary and highly desirable.

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